PB89-194047

Proceedings: Symposium on the Transfer and Utilization of Particulate Control Technology (7th). Volume 2

Radian Corp., Research Triangle Park, NC

Prepared for:

Environmental Protection Agency, Research Triangle Park, NC

May 89

U.S. Department of Commerce
National Technical Information Service
NTIS

PB89-194047

EPA/600/9-89/046b
May 1989

PROCEEDINGS: SEVENTH SYMPOSIUM ON THE TRANSFER AND UTILIZATION OF PARTICULATE CONTROL TECHNOLOGY
VOLUME 2

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U.S. DEPARTMENT OF COMMERCE
NATIONAL TECHNICAL INFORMATION SERVICE
SPRINGFIELD, VA. 22161

TECHNICAL REPORT DATA

(Please read Instructions on the reverse before completing)

1. REPORT NO. EPA/600/9-89/046b	2.	3. RECIPIENT'S ACCESSION NO.
4. TITLE AND SUBTITLE Proceedings: Seventh Symposium on the Transfer and Utilization of Particulate Control Technology, Volume 2		5. REPORT DATE May 1989 6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S) Susan R. Fields, Compiler		8. PERFORMING ORGANIZATION REPORT NO. EPRI GS-6208 (Volume 2)
9. PERFORMING ORGANIZATION NAME AND ADDRESS Radian Corporation P.O. Box 13000 Research Triangle Park, North Carolina 27711		10. PROGRAM ELEMENT NO. 11. CONTRACT/GRANT NO. NA (EPRI funded)
12. SPONSORING AGENCY NAME AND ADDRESS **EPA, Office of Research and Development** Air and Energy Engineering Research Laboratory Research Triangle Park, NC 27711		13. TYPE OF REPORT AND PERIOD COVERED Proceedings; 3/87 - 3/88 14. SPONSORING AGENCY CODE EPA/600/13

15. SUPPLEMENTARY NOTES AEERL project officer is Geddes Ramsey, Mail Drop 61, 919/541-7963. Symposium held in Nashville, TN, March 22-25, 1988.

16. ABSTRACT The proceedings document presentations from the seventh symposium on the transfer and utilization of particulate control technology, March 22-25, 1988, in Nashville, TN. Objectives of the symposium were to encourage the exchange of new knowledge in the particulate control field between designers, users, and researchers, and to stimulate new ideas that will lead to further technological improvements. Fabric filters (FFs) and electrostatic precipitators (ESPs) were the major topics, but novel concepts and advanced technologies were also explored. Volume 2 contains 30 papers covering particulate control for incinerators, full-scale FF studies, advanced SO2/particulate control studies, and FF pilot-scale studies.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS		b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group	
Pollution	Sulfur Dioxide	Pollution Control	13B	07B
Particles		Stationary Sources	14G	
Dust		Fabric Filters	11G	
Incinerators		Particulate		
Fabrics			11E	
Filtration			07D	

18. DISTRIBUTION STATEMENT **Release to Public**	19. SECURITY CLASS *(This Report)* Unclassified	21. NO. OF PAGES 511
	20. SECURITY CLASS *(This page)* Unclassified	22. PRICE

EPA Form 2220-1 (9-73)

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ABSTRACT

These Proceedings document the presentations from the Seventh Symposium on the Transfer and Utilization of Particulate Control Technology, held March 22-25, 1988, in Nashville, Tennessee. The Symposium was sponsored by the U. S. EPA's Air and Energy Engineering Research Laboratory in Research Triangle Park, North Carolina, and EPRI's Coal Combustion Systems Division, in Palo Alto, California.

The objectives of the Symposium were to encourage the exchange of new knowledge in the particulate control field among designers, users, and researchers, and to stimulate new ideas that will lead to further technological improvements in the future. Fabric filters and electrostatic precipitators were the major topics, but novel concepts and advanced technologies were also explored. Topics discussed in each day's sessions are listed below:

Tuesday, March 22

- Plenary Session
- FF: Particulate Control for Incinerators
- ESP: Dust Properties
- FF: Full-Scale Fabric Filter Studies I
- ESP: Power Supplies and Controls for Precipitators

Wednesday, March 23

- FF: Full-Scale Fabric Filter Studies II
- ESP: Advanced SO_2/Particulate Control Studies I
- FF: Advanced SO_2/Particulate Control Studies II
- ESP: Precipitator Performance Evaluation and Upgrading I

Thursday, March 24

- FF: Advanced SO_2/Particulate Control Studies III
- ESP: Precipitator Performance Evaluation and Upgrading II
- FF: Fabric Filter Pilot-Scale Studies I
- ESP: Precipitator Modelling I

Friday, March 25

- FF: Fabric Filter Pilot-Scale Studies II
- ESP: Precipitator Modelling II

Volume 1 of the Proceedings contains 33 papers presented in the Dust Properties, Power Supplies and Controls for Precipitators, Advanced SO_2/Particulate Control Studies I, Precipitator Performance Evaluation and Upgrading I and II, and Precipitator Modelling I and II Sessions. These papers are primarily concerned with electrostatic precipitators for particulate control.

Volume 2 of the Proceedings contains 30 papers presented in the Particulate Control for Incinerators, Full-Scale Fabric Filter Studies I and II, Advanced SO_2/Particulate Control Studies II and III, and Fabric Filter Pilot-Scale Studies I and II Sessions. The papers in this volume cover a broader range of topics, but the emphasis is on the use of fabric filters for particulate control.

ACKNOWLEDGMENT

The assistance of Joan Ross, Radian Corporation, in compiling these Proceedings is gratefully acknowledged.

PREFACE

The Seventh Symposium on the Transfer and Utilization of Particulate Control Technology was held in Nashville, Tennessee, March 22 through 25, 1988. This symposium, the third of its kind to be jointly sponsored by EPA and EPRI, was devoted to the discussion of particulate emissions control. The Symposium's objectives were to promote the transfer of results of particulate control research to potential users of technologies, and to provide a medium for exchange of ideas among researchers in the particulate control field.

The symposium proceedings contain 63 papers presented by EPA and other Federal and State agency representatives, EPRI staff members, representatives of utility companies, equipment and process suppliers, university representatives, and research and development companies. Additionally, individuals from Australia, Canada, Denmark, Holland, Sweden, and Switzerland presented information on worldwide technological developments. Electrostatic precipitators and fabric filters were the major topics discussed during the Symposium. Other topics presented included system operation and maintenance, and integrated control processes.

Symposium cochairmen were Ralph Altman, Project Manager in the Air Quality Control Program of the EPRI's Coal Combustion Systems Division; and Geddes Ramsey, Project Officer in the Air Toxics Control Branch of the EPA's Air and Energy Engineering Research Laboratory. Welcoming remarks made for EPA by Geddes Ramsey, and for EPRI by Dr. George Offen, Program Manager in the Air Quality Control Program of EPRI's Coal Combustion Systems Division.

The Seventh Symposium broke new ground in presenting a panel discussion of particulate control issues at the beginning of the Symposium. Participants in the plenary session Panel Discussion on Particulate Control Issues were **Charles Altin**, Project Manager, Ebasco Services, Inc., **A. A. Arstikatis**, Client Manager, Combustion Engineering, **Peter Bibbo**, Vice President, Sales and Marketing, Research-Cottrell, **Gary Greiner**, Vice President, ETS, Inc., **Gerald Hollinden**, Branch Chief, Tennessee Valley Authority, **Richard McRanie**, Director of Utility Services, Kilkelly Environmental Associates, **Grady Nichols**, Director of Environmental Sciences Research, Southern Research Institute, **David Sloat**, Senior Engineer, Sargent and Lundy, **John Walton**, Deputy Director, Tennessee State Department of Health and Environment Division of Air Pollution Control, and **Jeffrey Williams**, Environmental Analyst, Rochester Gas and Electric. A summary of the discussion is included in Volume 1 of these Proceedings.

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Introduction

Emissions from sewage sludge incinerators are evaluated when the system is placed on-line and tested and during subsequent periodic emissions tests. Between tests, system operation and maintenance sometimes fails to meet the intent of the original engineer, contractor and regulatory body. As such, it is imperative that each incineration facility be designed and operated in such a manner that the original philosophy can and will be achieved by the owner. The largest single emission from most incineration facilities is dry particulate, as determined by EPA method 5 tests. This document will address the various types of incinerator emissions with particular emphasis on particulates. It will then specifically identify some design/operating characteristics which impact particulate emissions. Data is incorporated from facilities designed and constructed by Nichols Engineering and Research Corporation. This data provides representative results of many facilities tested under current guidelines during the past ten years.

The incineration of sewage sludge is most commonly accomplished in two types of furnace configurations. The multiple hearth is the most common, followed by the fluidized bed furnace. Other types of equipment have been utilized during the past fifty years, but none have proven to be as successful or popular as these configurations. Both can and do incinerate sewage sludge solids with a minimum of emissions. Each has it's own operating quirks which impact emissions types and rates. It is the purpose of this paper to identify the characteristics of emissions from these systems burning sludge and to provide some insight on the system operating parameters and physical mechanisms that may be used to minimize those emissions.

Multiple Hearth Incinerator

The multiple hearth furnace (MHF) is unique in many ways in the process industry. The furnace is a vertical cylindrical shell with refractory hearths separating multiple refractory lined chambers. The cross section of a typical multiple hearth incinerator is shown in Figure 1. Individual burner and combustion air controls can make each chamber (hearth) somewhat independent of the hearths above and below with regard to temperature and atmosphere. Temperature gradients within the furnace can be varied to suit the particular feedstock. The sludge feed is introduced into one of the upper hearths and is mechanically moved across each hearth by flat alloy "teeth" attached to the rabble arms. The material is pushed to the edge of the hearth to the "drop hole" to drop to the hearth below. After drying and combustion of volatiles, fixed carbon burns in the bed. The remaining ash eventually falls from the furnace discharge.

The physical arrangement of the MHF provides the advantage of countercurrent flow of thecombustion gases and sludge solids.

The end result is a low furnace outlet temperature with a relatively high (1400-1800°F) combustion temperature. This arrangement has proven to be very fuel efficient. The low furnace outlet temperature yields a low fuel consumption per ton of sludge. This remains an advantage only until secondary combustion (afterburning) is required for an additional reduction in organic vapor emissions. Once this condition occurs, the fluidized Bed (FB) system has a lower fuel consumption per ton of sludge incinerated. A typical FB system operates with a lower percentage of excess air due to the high heat and mass transfer rates in the bed.

The combustion process in the multiple hearth is unique because of the stacked hearth design. The process of incineration in any furnace requires several things to occur. In the MHF these things are easily observable, while in a FB they appear to occur simultaneously.

o First, the sludge warms from ambient temperature to a temperature where significant drying begins. This usually occurs at about 150 to 180°F. At that point the temperature of the particle (which may vary from pea to baseball size) remains fairly constant, perhaps warming slightly.

o Once the outside surface of the particle reaches dryness, the thin surface layer begins to heat to a temperature of 200-400°F. The mean particle temperature is still only about 180°F, and is still drying, but light organics are being driven from the surface simultaneously.

o Sludge will normally begin to burn at about 50% solids at that average temp. because of the substantial temperature gradient from the outside to the inside of the particle. Burning usually is completed quickly, requiring only one to two hearths. The remainder of the moisture is driven off as the temperature gradient rises toward the center of the particle.

o Combustion of the fixed carbon in the ash occupies the rest of the furnace. This process begins quickly and then slows as the amount of available carbon is reduced and the remaining ash blinds the surfaces of the pores in the ash. A great deal of carbon monoxide is formed in the bed in this zone of the furnace regardless of the amount of oxygen available in gas phase. This carbon monoxide burns in the gas phase and in the combustion zone above. In most cases, combustion of the fixed carbon does not reach completion and the reaction simply stops for lack of sufficient temperature as the reaction slows or locally available oxidant (the oxidant is carbon dioxide which diffuses into the bed to react with carbon and form carbon monoxide.)

The typical multiple hearth incinerator operates in an oxidizing atmosphere, may or may not have an air to air heat exchanger or boiler for heat recovery and relies on a venturi scrubber for particulate removal in accordance with USEPA guidelines.

Figure 2 shows a process schematic for a typical MHF incineration system. This configuration is well suited for the average treatment plant for operational purposes. However, from the standpoint of emissions control, much better methods of both operation and emissions capture are available, especially as many facilities are incinerating sludges with a myriad of industrial components. Costs for alternative methods are higher and operation more sophisticated. For this reason, the transition away from venturi scrubbers has not yet been made.

The venturi scrubber normally operates with a pressure drop of from 18 to 40 inches of water column. Venturi water is sometimes recycled, however, quench water upstream of the venturi is always as clean as possible. It has been found that the evaporation of dirty water in the quencher can lead to the release of very fine particles into the gas phase from the evaporation to dryness of some of the water droplets in the quencher. The ideal situation would be to use potable water to quench the combustion gases before reaching the venturi. Again, the cost of potable or clean process water usually precludes it's use.

Selection of actual operating pressure drop is made based upon available data for sludge ash characteristics. If none is available, a design of 25 inches is common. The impingement tray section of the scrubber is normally provided to subcool the exhaust gases thereby reducing plume and the "visibility" of the facility. The use of wastewater treatment plant effluent as subcooling water occasionally adds odors to the exhaust gas which can be problems downwind.

Overloading a MHF incineration system will cause emissions in several ways.

o Increased elutriated particulates will increase loadings to the venturi scrubber. A larger quantity of fine particles will increase emissions from the air pollution control system.

o The problem of locally increased oxygen demand will result in soot formation even though oxygen concentration at the furnace outlet is satisfactory.

o A continued increase in feedrate will result in unstable operation and burning rate. The rate will initially be depressed by the additional wet feed. Later, the burning rate will increase far above that equivalent to the current feed rate. Oxygen demand will far exceed that available and a period of several minutes of operation at an oxygen starved condition will follow. Burning rates under these conditions tend to follow a sinusoidal pattern gradually approaching a new equilibrium after 30 to 60 minutes if new process disruptions are avoided.

As design and operation of the typical sludge incinerator becomes more sophisticated in the future, the multiple hearth may again gain advantage. It's ability to vary operating temperatures, and especially atmosphere, from oxidizing to reducing, may enable specific process design for the municipality which is faced with a particular metal(s) contaminating their sludge feed. As in the mining industry, the process can be tailored to maintain the metal in a form which will remain in the ash, or it may be fumed and driven into the gas phase to be captured by a high energy scrubber, baghouse or precipitator.

Fluidized Bed

The development and use of the fluidized bed (FB) incinerator for sewage sludges came as a later development to the industry. In the FB system, air is pressurized to 5 psig and introduced into the bottom chamber of the refractory lined steel vessel. A burner may be installed in this chamber or above the expanded bed for operation during startup to preheat the system. The high pressure air travels upward through a perforated air distribution plate or refractory arch.

Typical pressure drops across the air distribution plate are 15 to 20 inches of water column. The superficial upward velocity of the gas ranges from 3 to more than 5 feet per second, sufficient to fluidized the bed media. The most common choice of bed media is silica sand, but slag products such as "Black Beauty" are also used for their increased density and reduced attrition rates during operation. The bed material selected is of a higher density and larger diameter than sludge ash. Most bed materials range between 10 and 40 mesh size depending upon superficial velocity and material density.

Sludge can be fed from above via a pipe and positive displacement pump or through the side of the vessel with a conveyor. Primary combustion takes place in the bed unless the system is operating at or above design capacity. The freeboard (volume above the expanded bed) allows most entrained particles to fall back into the bed.

Many units provide a slightly increased diameter in the freeboard area to decrease superficial velocity. This expanded area reduces superficial velocity and decreases attrition of bed material by reducing carryout of the larger diameter bed material.

All ash generated by the combustion of the sludge passes from the incinerator with the combustion gases. Air pollution control is normally achieved with a recirculating venturi scrubber. Most FB incineration system scrubbers are equipped with impingement coolers to subcool gases to a range of 90 to 120° F. A process schematic of a typical fluidized bed incineration system is shown in Figure 3.

Overloading the unit can increase emissions in several ways.

- Particulate (ash) carryout will increase to the venturi scrubber.

- Reduced oxygen concentration will result in an increase in emissions of hydrocarbons, carbonyls and carbon monoxide. Further decrease in available air for combustion will result in a decrease in bed/furnace temperature and resultant emissions of tars, oils and greases in vapor form. The FB system will operate at oxygen concentrations of 30 to 50% excess air, while a MHF requires 75 to 150% excess air for success.

Characterization of Emissions

From the perspective of the incinerator designer, constructor or operator, emissions from sludge incinerators fall into the following categories.

- Dry Particulate (fly ash)
- Condensable Organics, Oils and Greases
- Mercury
- other Heavy Metals
- Acid Gases

In most installations, control technologies are implemented only for dry particulate fly ash, as the quantities of other emissions are either not regulated or not present in quantities sufficient to require control. In recent years, however, incinerator operators are finding themselves under increasing regulatory and public scrutiny. Industrial waste streams are becoming a larger percentage of the total waste processed in many municipal wastewater treatment facilities. These streams range from innocuous to major contributors to the plants incinerator stack emissions. Major problems in emissions control have resulted from industrial waste stream contributions to facilities which were designed to process municipal wastewater only.

Dry Particulate:

Fly ash from the normal incineration process contains ash from the municipal sludges as well as ash resulting from the incineration of any industrial sludges which may be present. These industrial components may include heavy metals, which are discussed below, and traces of other unusual materials.

A typical incinerator fly ash is light brown in color, with a particle size ranging from one micron to more than forty microns in diameter. A small fraction of most ash residues is less than one micron in diameter. Significant quantities of material in this size range are unusual.

Since the furnace ash has been burned in a slightly reducing to slightly oxidizing atmosphere, it is composed of metallic oxides (lime, iron oxides, silica) and small quantities of base metals. The heavy metal constituents of the ash is usually small. However, these constituents may tend to concentrate in the fly ash. Particle diameter of the fume formed by the reduction and/or vaporization of metals is often much less than one micron in diameter and thus can be impossible to effectively remove with a venturi impingement scrubbing system.

Many studies have been made to establish that the current accepted standards of design for control equipment can meet federal and state codes for incinerator emissions. The conclusion of these reports has been that the systems currently in use and under construction with venturi and equivalent scrubbing systems can meet code requirements when operated properly. However, there is one subgroup that may not be able to meet these rules. That group is the subset of sludge handling facilities that produces sludge filter or centrifuge cakes of solids contents of less than 24% solids. Note that the lower the solids content, the greater the likelihood of a problem.

Certain types of sludge and sludge treatment result in filter or centrifuge cakes of high moisture content (between 14 and 20 percent solids.) The particulate code which these feedstocks must meet when incinerated is 1.3 pounds per ton of dry solids incinerated. Feedstocks of this low solids content are generally not autogenous. That means that they normally require the addition of auxiliary fuel to burn. If the feed has a low combustible content, as many digested sludges do, then even more fuel must be added to the wet sludge to evaporate the water and burn the remaining solids.

The additional fuel carries with it combustion and excess air. This gas stream entrains a significant amount of particulate which the scrubber must remove. The emissions codes for particulate emissions from sludge incinerators is currently 1.3 pounds per ton of dry sludge feed. These additional combustion products result in an increase in the quantity of emissions with no allowance in the codes for additional allowable emissions.

Specific instances of emissions from facilities such as this are listed in the Table 3, while Table 2 lists results from plants which incinerate drier feedstocks. Note the comparison between the actual results and the code limitations. These results are representative of the bulk of facilities constructed and tested by Nichols over the past twelve years. While not every plant with wet feedstock has a problem, the trend indicated by the tables is more the rule than the exception.

This problem is further complicated by an operating characteristic of the MHF. In the incineration process, combustion products are passed across each hearth before rising through the drop hole to the hearth above (see Figure 1).

Early research by Nichols, as well as physical observation of many facilities has shown that elutriation of the ash occurs as the ash is falling from hearth to hearth through the drop hole. Above about 15 to 20 acfm of velocity up through the dropholes at and below the burning zone dramatically increase the rate at which particulates are carried out. This is confirmed by visual observation of the increase in particle carryout to the scrubber.

The FB carries out all of the ash resulting from the incineration process. From Table 1, it can be seen that FB systems tend to emit smaller quantities per ton of feed than similar MHF counterparts (Table 2), in spite of much greater loadings to the scrubber.

One unusual contaminant which has appeared on several occasions during the past few years is submicron silica. Uniform silica particles at less than 0.5 micron diameter are nearly impossible to catch in incinerator emissions control systems.

A specific case of submicron silica in the feedstock of a multiple hearth sludge incinerator occurred several years ago. Initial emissions tests resulted in rates of 3 to 4 pounds of particulate emissions per ton of dry feed. Further tests run at 18 and 40 inches of water column yielded virtually identical emissions rates. An attempt was made to determine a particle size distribution. An Anderson Cascade Impactor with a modified method 5 train was utilized to determine particle size distribution of emissions in the stack. Results indicated that 50 to 60 percent of the particulate captured were less than 0.5 micron.

The particles were identified using scanning electron microscopy as silicon dioxide particles, uniform in size between 0.2 and 0.3 microns. The particles appeared to carry a negative charge sufficient to make scrubbing impossible with the standard venturi scrubber provided. Electron microscopic examination of larger size cut filters showed that most of the particulate captured on all of filters were agglomerations of the same 0.2 micron particles.

Resolution of this kind of problem requires either that the contaminant source be identified and eliminated, or that the pollution control system be upgraded substantially to handle the fine particles.

Condensable Organics:

As a discrete sludge particle is dried in the incinerator, greases, oils and long chain organics volatilize to the gas phase. In a FB incinerator, these materials are immediately broken down and burned, because the operating temperature of this zone of the system will normally be in excess of 1400^{O}F. In a multiple hearth incinerator, the gas phase temperature in this zone may range as low as 600^{O}. Locally lower temperatures may also occur.

The temperatures encountered in the MHF drying zone are insufficient to ensure combustion of condensable organics, unless an afterburner or zero hearth is incorporated into the design. Locally poor mixing of gases can result in poor combustion which will also contribute to this class of emissions. As the gases are scrubbed, these emissions may condense on dry particulates, or form droplets. The presence of an oily surface on a particle will make it much more difficult to capture in a wet scrubber.

Instances of poor combustion and grease and oil emissions are very common in the day to day operation of many MHF incinerators. In performance evaluation, these combustibles often show up as dry particulate on the filter pad or in the acetone probe wash of the method 5 train. A high emission rate for this type of system upon testing can be checked first for hexane extractable condensables from the filter pad. Greases and oils will be identified in this way and can be eliminated by proper operation of the system.

Several techniques that can be used to reduce condensable emissions if no afterburner is available are:

o Increase the center shaft rotation speed to prevent the surface of the sludge particles in the furnace from becoming too hot between passes of the rabble arms. Additional agitation of the bed on the hearth will allow the particle to become more uniformly dry. The organics with their correspondingly higher boiling temperatures will remain on the particle until combustion begins.

o Separate the greases and oils from the sludge feed and inject separately into the burning zone of the furnace. If this method is chosen, glass lined piping and heat tracing is recommended, as well as preheating of the grease and oil feed. Note also that the incineration of sewage skimmings, greases and oils tends to result in the generation of a greater quantity of fine particulate carryout than an equivalent quantity of sludge solids.

o Ensure that the sludge solids are properly distributed on the surface of the hearths to eliminate locally reducing atmospheres. Poorly distributed material on the upper, cooler hearths may lie in the bed below the reach of the rabble teeth and smolder, creating substantial condensable and smoke emissions.

o Use burner air from unused burners as combustion air in high heat release areas. The additional turbulence will increase mixing, improve combustion heat release and combustion efficiency for the sludge solids.

Mercury Emissions:

Mercury emissions have proven to be very difficult to control by conventional methods on sludge incinerators. Collection efficiencies can range from below 50 to 96%. In most cases, however, the quantities of mercury under consideration are small when compared to the allowable limitations on total mercury emissions.

Metals Emissions:

The quantities and character of the metals in sludge feeds are almost always a function of the industrial contribution to the treatment plant influent. Substantial information is currently available in the literature concerning the emissions of metals from specific facilities. (See ref. 2, 3, 4, 5 and 6.) The metals of greatest concern seem to be lead and cadmium. These metals are most likely to be found in municipal sludges and have a significant tendency to vaporize at normal combustion temperatures. This circumstance occurs in both the MHF and FB systems, as the ash temperatures normally reach about 1400 degrees F in both systems.

In tests conducted by the USEPA (ref. 2) it was found that Cadmium (Cd) and Lead (Pb) concentrate in the particulate emissions by factors of up to 10 for Cd and 25 for Pb as operating temperatures of a fluid bed increase from 1200 to 1750°F. Work performed by Nichols Engineering in the late seventies indicated similar increases in emissions rates vs. temperatures.

It may be possible, with proper design and consistent feedstock quality, to operate a system in such a manner to significantly reduce the emissions of heavy metals. Knowledge of the metals involved is of paramount importance. Once identified, each metal can be addressed individually to determine if the mechanism for emission is physical, chemical or both.

These mechanisms include:

- Physical carryout of fine particulate containing the metal oxide or other salt.
- Reduction to elemental metal in the furnace with subsequent volatilization to metal vapor.
- Volatilization of the metal oxide or other salt directly into the gas phase with no chemical change.

It may be logically assumed that a reduction in metals volatilization can be achieved by reducing the temperature to which the ash is exposed. However, in reality, this is not practical in a normal incinerator.

The fluidized bed incinerator continuously operates at temperatures in the range of 1350 to 1700 degrees F with a slightly oxidizing atmosphere (30-50% excess air.) Substantial reductions in operating temperatures cannot practically be made in this case. A decrease in temperature would substantially reduce system capacity, while an increase (noting that even 1750 is quite high) risks slagging of the bed and subsequent extended shutdown.

Ash Temperature in the MHF is determined by the amount of fixed carbon generated in the burning zone of the furnace and by the degree to which the sludge ash hinders diffusion of CO_2 into the ash bed to burn this fixed carbon. A more rapid combustion of fixed carbon results in a higher ash temperature and subsequent increase in metal volatilization. Typical ash temperatures during the fixed carbon burning process ranges from 1200 to more than 1700^{o}F.

The multiple hearth, however, can limit ash temperature by limiting the number of hearths in the furnace or by other methods. Note that when burning combustibles in a MHF, the actual sludge temperature does not rise above about 500 degrees F until all of the volatiles have been driven from the bed. At that point, only ash and fixed carbon remain.

Multiple hearth systems have been designed to operate as incinerators and also as pyrolyzers. In the pyrolysis mode, the furnace is constructed somewhat differently, to provide for a tighter control on combustion air infiltration. The furnace shell is of welded rather than bolted construction and the top and bottom of the furnace incorporates water seals. This serves to allow the operator to introduce only enough air to burn a portion of the combustibles. Lack of air in the fixed carbon burning zone of the furnace will limit the temperature to which the ash can climb. The disadvantage of this method is the loss of the heating value of the fixed carbon, which can be as much as one third of the total heat content of the feeds.

However, the MHF Pyrolyzer operates at much lower excess air conditions than a normal incinerator and thus requires less fuel than an incinerator in virtually all cases where feedstock solids content exceeds 28%. Pyrolytic incineration of sludge is a method that should be considered whenever metals concentrations are high. The physical/chemical mechanism of volatilization must be reviewed for the specific metals in the proposed feedstock to verify that pyrolysis will result in a substantial reduction in emissions.

Acid Gases:

All sewage solids contain some Sulfur and Chlorine. In addition, many facilities still condition their solids for dewatering using ferric chloride and lime rather than polymer. The chloride contribution from ferric chloride can be substantial.

Much of the chlorine from ferric chloride results in hydrochloric acid formation. Wet venturi scrubbing of emissions by most facilities with perforated tray subcoolers allows sufficient contact for the removal of most of the HCl from the stack gases. The risks attendant to high chlorine levels in the feed lie not with the stack emissions but with possible severe corrosion to the upper sections of the furnace and the emissions control equipment.

Materials of construction for these systems range from 316L stainless to Hastelloy or Inconel alloys in some applications.

In the case of sulfur, a typical feed contains from 1 to 4 percent sulfur on a combustible basis. Only a fraction of this results in sulfur oxide emissions. That percentage will depend greatly upon the remaining constituents in the feeds. The use of lime in both water and wastewater treatment plants for clarification results in a lime sludge which often ends up at the wastewater treatment facility for disposal. Thus, even when no lime is added at the final stages of solids handling, there is a substantial probability that it will be present and reduce the concentration of sulfur oxides emitted.

Conclusions

- Both Multiple Hearth and Fluidized Bed Sludge Incineration Systems can continue to meet federal emissions codes for particulates when burning municipal sewage wastes at solids contents greater than 20%.

- When sludge feedstock solids contents fall below 20% the incinerator may have difficulty meeting existing federal emissions codes, especially in multiple hearth systems.

- The growing fraction of industrial wastes in municipal treatment plant wastewater influent has dramatically increased the problem of heavy metal emissions from sludge incinerators.

- As increases in metals emissions continue, and emissions regulations for metals are promulgated, sludge incinerator operators will be forced to make system modifications to achieve compliance. Technologies exist today for the removal of these metal contaminants. Utilizing today's technologies in the areas of fabric filtration and electrostatic precipitation, metal removal, and perhaps recovery, is the next step in the evolution of the incineration of sludge.

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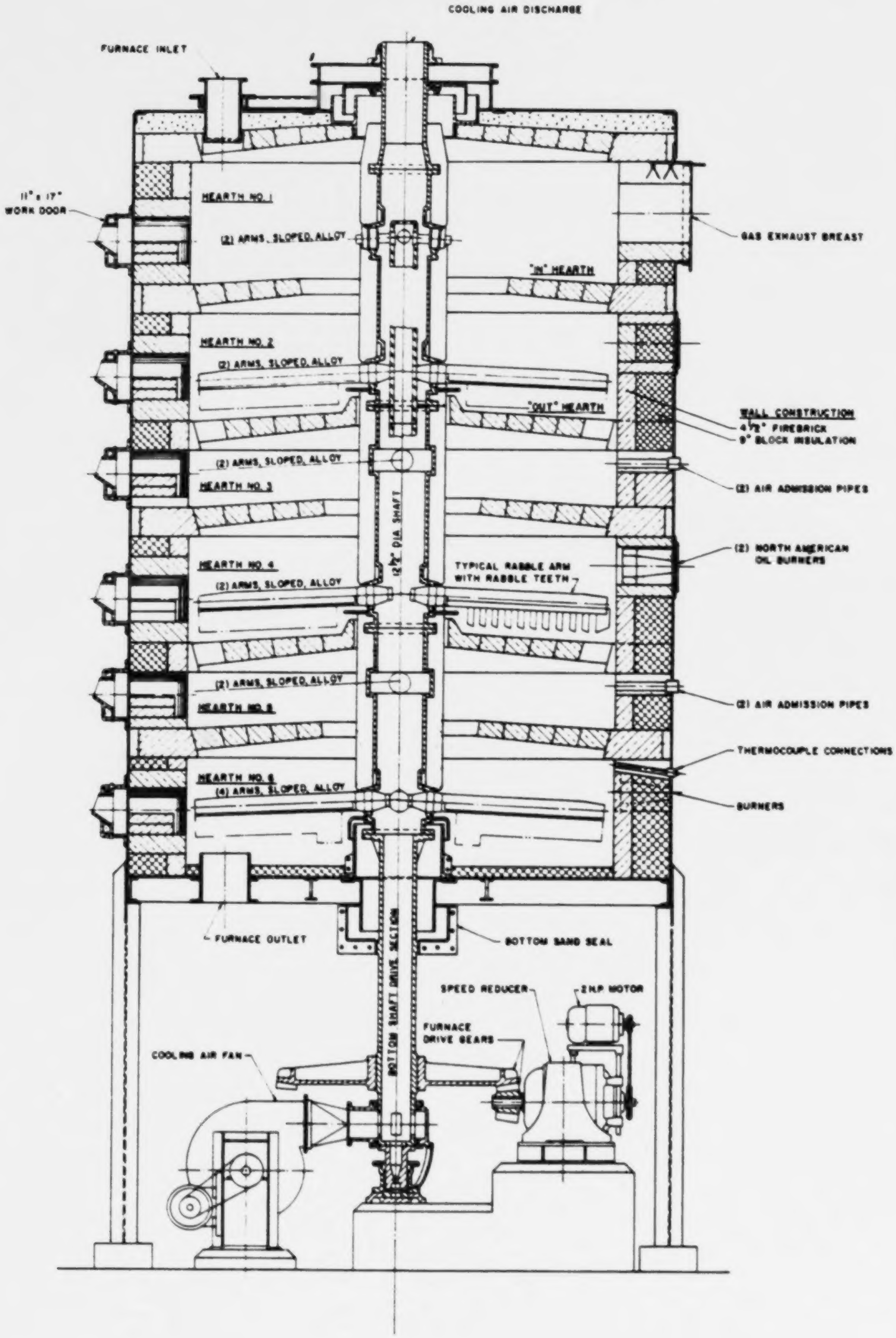
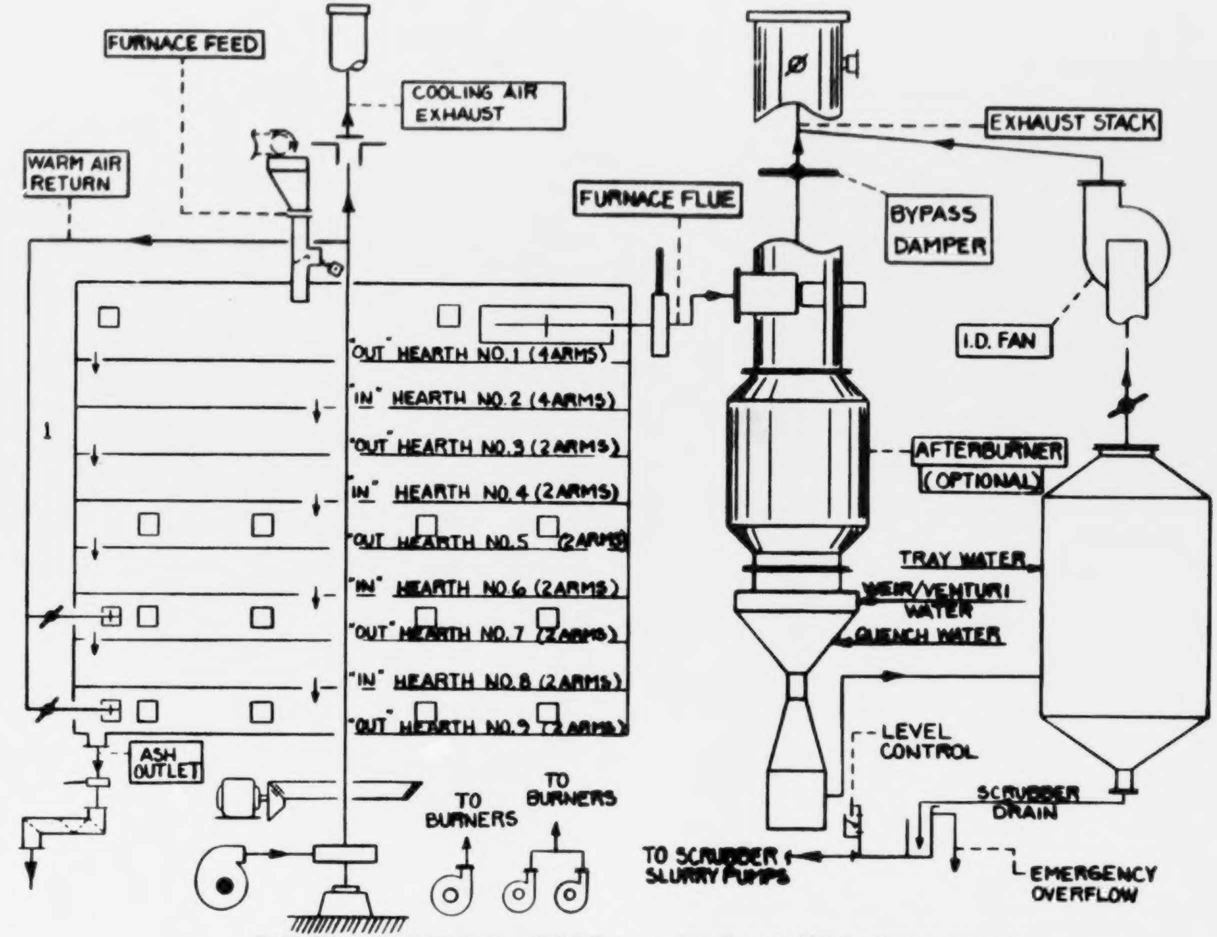
The work described in this paper was not funded by the U.S. Environmental Protection Agency. The contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

Figure 1

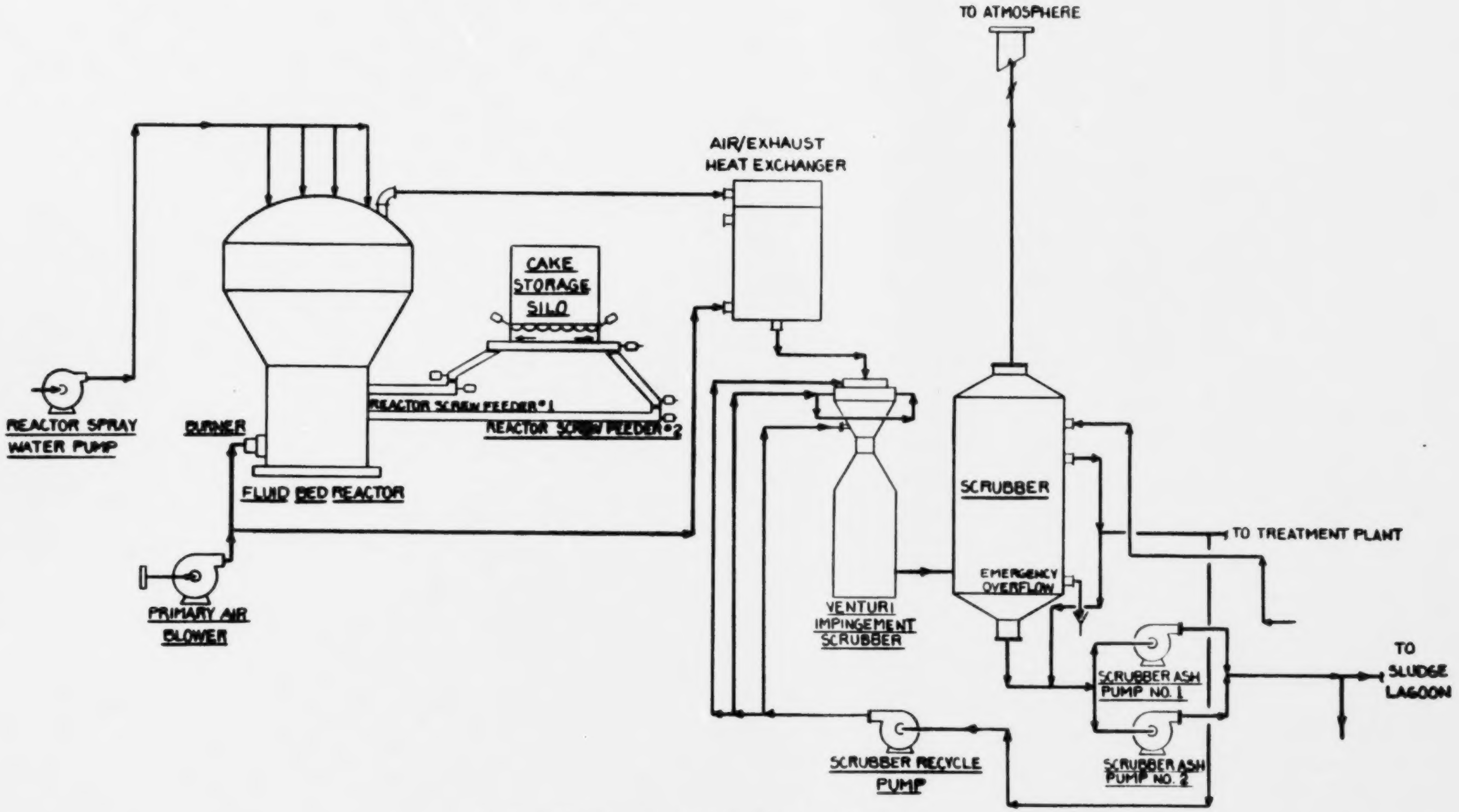
Multiple Hearth Incinerator Cross Section





TYPICAL MULTIPLE HEARTH SLUDGE INCINERATION SYSTEM

FIGURE 2



FLUID BED SLUDGE INCINERATION SYSTEM
FIGURE 3

Table No. 1

Particulate Emissions Results from Fluidized Bed incinerators utilizing Venturi Impingement Scrubbing Systems.

Facility	Type of Incinerator	Scrubber Config.	Feed Solids Concentration	Feedrate, Tons/Hour		Gas Flow	Flue Gas Flow	Emissions Rates		Federal Code
				WET	DRY	SCFM Dry	SCFM/TON Feed	Grains/SDCF	#/Ton Dry Fe	#/Ton Dry Feed
1	Fluidized Bed	Venturi/Impingement	36.1%	1.483	0.54	1392	2600	0.0152	0.318	1.3
2	Fluidized Bed	Venturi/Impingement	40.7%	1.421	0.58	1214	2099	0.0096	0.190	1.3
3	Fluidized Bed	Venturi/Impingement	42.1%	1.554	0.65	1418	2167	0.0167	0.306	1.3
4	Fluidized Bed	Venturi/Impingement	31.4%	4.231	1.33	4166	3141	0.0035	0.097	1.3
5	Fluidized Bed	Venturi/Impingement	34.0%	4.163	1.42	4338	3065	0.0020	0.054	1.3
6	Fluidized Bed	Venturi/Impingement	34.0%	4.326	1.47	4400	2991	0.0046	0.120	1.3
7	Fluidized Bed	Venturi/Impingement	28.0%					0.0004		1.3
8	Fluidized Bed	Venturi/Impingement	28.0%					0.0005		1.3
9	Fluidized Bed	Venturi/Impingement	28.0%					0.0007		1.3
	AVERAGE	RESULTS	33.6%	2.863	1.00	2821	2677	0.0059	0.181	

Table No. 2

Particulate Emissions from Multiple Hearth Incineration Systems Utilizing Venturi Impingement Scrubbing Systems

Subset A, High Solids Content Feedstocks

Facility Test	Type of Incinerator	Scrubber Config.	Feed Solids Concentration	Feedrate, Tons/Hour WET	DRY	Gas Flow SCFM Dry	Flue Gas Flow SCFM/TON Feed	Emissions Rates Grains/SDCF	#/Ton Dry Fe	Federal Code #/Ton Dry Feed
10	Multiple Hearth	Venturi/Impingement	28.3%	6.260	1.77	15248	8607	0.0015	0.113	1.3
11	Multiple Hearth	Venturi/Impingement	27.2%	6.650	1.81	15923	8806	0.0035	0.260	1.3
12	Multiple Hearth	Venturi/Impingement	29.3%	6.650	1.95	16258	8347	0.0052	0.369	1.3
13	Multiple Hearth	Venturi/Impingement	31.7%	6.225	1.97	9412	4771	0.0102	0.420	1.3
14	Multiple Hearth	Venturi/Impingement	31.3%	6.610	2.07	9765	4727	0.0111	0.450	1.3
15	Multiple Hearth	Venturi/Impingement	33.9%	6.920	2.35	9349	3985	0.0099	0.340	1.3
16	Multiple Hearth	Venturi/Impingement	26.1%	4.483	1.17	9580	8188	0.0182	1.280	1.3
17	Multiple Hearth	Venturi/Impingement	28.7%	4.390	1.26	9520	7556	0.0145	0.930	1.3
18	Multiple Hearth	Venturi/Impingement	27.0%	4.444	1.20	9510	7925	0.0150	1.020	1.3
19	Multiple Hearth	Venturi/Impingement	29.9%	6.397	1.91	12931	6770	0.0028	0.162	1.3
20	Multiple Hearth	Venturi/Impingement	31.7%	6.317	2.00	14295	7139	0.0020	0.125	1.3
21	Multiple Hearth	Venturi/Impingement	29.7%	5.430	1.61	13470	8366	0.0039	0.280	1.3
22	Multiple Hearth	Venturi/Impingement	36.5%	7.010	2.56	15309	5991	0.0094	0.482	1.3
23	Multiple Hearth	Venturi/Impingement	33.2%	5.830	1.94	15296	7895	0.0096	0.661	1.3
24	Multiple Hearth	Venturi/Impingement	34.0%	6.160	2.09	14851	7099	0.0109	0.664	1.3
	AVERAGE RESULTS		30.6%	5.985	1.84	12714	7079	0.0085	0.504	

Table No. 3

Particulate Emissions from Multiple Hearth Incinerators Utilizing Venturi Impingement Scrubbing Systems

Subset B: Low Solids Content Feedstocks

Facility Test	Type of Incinerator	Scrubber Config.	Feed Solids Concentration	Feedrate, Tons/Hour		Gas Flow SCFM Dry	Flue Gas Flow SCFM/TON Feed	Emissions Rates		Federal Code #/Ton Dry Feed
				WET	DRY			Grains/SDCF	#/Ton Dry Fe	
25	Multiple Hearth	Venturi/Impingement	23.6%	7.430	1.75	21500	12286	0.0231	2.430	1.3
26	Multiple Hearth	Venturi/Impingement	24.3%	7.440	1.81	21800	12044	0.0399	4.120	1.3
27	Multiple Hearth	Venturi/Impingement	23.0%	7.910	1.82	21900	12033	0.0295	3.040	1.3
28	Multiple Hearth	Venturi/Impingement	13.9%	4.347	0.60	9430	15607	0.0166	2.218	1.3
29	Multiple Hearth	Venturi/Impingement	13.7%	4.292	0.59	9640	16395	0.0194	2.720	1.3
30	Multiple Hearth	Venturi/Impingement	15.0%	4.600	0.69	9380	13569	0.0207	2.400	1.3
31	Multiple Hearth	Venturi/Impingement	18.0%	2.481	0.45	10140	22701	0.0076	1.507	1.3
	AVERAGE RESULTS		18.8%	5.500	1.10	14827	14948	0.0224	2.634	

TOTAL SCRUBBING OF INCINERATION AIR POLLUTANTS --

THE EMCOTEK APPROACH

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ABSTRACT

A new type of wet scrubber providing baghouse-equivalent particulate removal has been developed and installed in a hospital infectious waste burning facility. This paper presents compliance test data obtained at this central-burn facility located in California. The test data includes both acid gas and particulate emissions based on operation of two separate flue gas cleaning systems. One system is a pulsejet baghouse equipped with pneumatic lime injection. The other system is the EMCOTEK Rotary Atomizing^{TM} scrubber which uses caustic reagent. Both systems were tested downstream from an Inconnel quench chamber.

At bio-hazardous waste burning rates which are nearly identical, the Rotary Atomizing^{TM} scrubber produced average particulate emissions of 0.022 gr/DSCF, corr. 7% O_2, approximately 60% less than the pulsejet baghouse system. The wet scrubber also produced 94% less organics and condensibles, such as dioxins and furans, compared with the baghouse system. The Rotary Atomizing^{TM} scrubber also produced 85% less acid gas emissions compared with the pneumatic lime-injected baghouse. The test results show that the wet scrubbing system using an EMCOTEK Rotary Atomizing^{TM} scrubber provides superior flue gas cleaning performance compared with a dry lime-injected baghouse system.

INTRODUCTION

A major concern in the design and operation of bio-hazardous waste incineration systems in the United States is control of flue gas emissions. At the Sixth Symposium (New Orleans, 1986), EMCOTEK discussed a novel and highly efficient liquid spray curtain impaction scrubber (1). The EMCOTEK Rotary Atomizing^{TM} scrubber has the significant advantages of removing both submicron particulate and acid gases, in one step, at low pressure drop, and at much lower temperature than a baghouse or ESP. The low temperature substantially reduces emission of organics such as dioxins and furans. This paper presents actual test data comparing the Rotary Atomizing^{TM} scrubber with a lime-fed baghouse, both in a commercially-operating bio-hazardous waste incinertion facility.

Bio-hazardous waste has a heating value of about 10,000 BTU/lb, compared with 4500 BTU/lb for typical municipal waste. It also contains a larger percentage of chlorinated and sulfonated plastics, which generate substantial levels (200 to 1000ppm) of hydrochloric acid and sulfur dioxide in the flue gases. Traces of incombustible oxides in the typical flyash generally contain heavy metals such as mercury, chrome, lead, cadmium or zinc. These escape as metal chloride and oxychloride vapor except for mercury metal, which has a high vapor presure in the metallic state. Incomplete combustion of organic matter in the secondary combustion chamber can allow traces of radicals called PICs to reach the scrubbing system. PICs include dioxins and furans, which are recombination products identified in combustion products, generally at the part per trillion level.

COMMERCIAL TEST FACILITY

The existing incinerator is a typical, modular, two chamber unit located in California. The incinerator has a design capacity of about 800 lb/hr of boxed bio-hazardous waste, operating 24 hr/day. The incinerator is normally operated continuously, and is loaded with hospital and veterinary infectious waste via a ram feeder. During incineration, the combustibles move down the primary incineration chamber by the action of new boxes being loaded by the ram. Following combustion, the ash is removed by the intermittent motion of a large chain conveyer belt. The conveyor transfers hot ash to a dumpster every 1 to 3 hours. The bottom ash differs from the flyash in that it has been mostly depleted in volatile heavy metals.

A secondary, post combustion chamber (or afterburner) follows the primary chamber. There are natural gas burners in this chamber. These burners allow temperature control during startup and operation, thereby completing the combustion of organic vapors. In this chamber, the flue gas temperature is controlled at 2000±200°F. The flue gases leaving the chamber pass through a ceramic-lined duct directly into a quench tank.

This quench tank, constructed of Inconel 625, is fitted with internal spray nozzles through which caustic soda is introduced via recirculation of a caustic brine solution. When the lime-fed baghouse is in use, just sufficient caustic brine solution is added to cool (via water evaporation) the 2000°F flue gases to 400°F, the normal operating temperature of the baghouse. During operation of the Rotary Atomizing^{TM} scrubber, excess caustic and byproduct soluble salts are recirculated through a larger spray nozzle to allow quenching of the incinerator gases to complete saturation with moisture, which cools the gases to the range of 155 to 170°F. The installation of the Rotary Atomizing^{TM} scrubber is shown in Figure 1, details of the Rotary Atomizer are shown in Figure 2, and the installation of the lime-injected pulsejet baghouse system are provided in Figure 3.

BAGHOUSE PERFORMANCE

Operation of the baghouse at 400°F over several months of incineration of bio-hazardous waste revealed a number of operating problems. The most urgent problems fell into the following categories:

a. Newly installed bags worked well, but incinerator throughput deteriorated rapidly to less than 50% of design, because of corrosion, bag plugging, and resulting loss of draft to the incinerator.

b. Corrosion in the mild steel duct to the steel baghouse and the baghouse itself was severe. This is attributed to the moisture, and to often very high levels of HCl gas in the entering gases, caused by difficulties maintaining dry lime feed into the airstream.

c. Particulate emissions were between 0.035 and 0.068 grains/DSCF, corrected to 7% O_2, (see Table 1A) compared with a compliance limit of 0.08 gr/DSCF, corrected to 7% O^2. HCl acid emissions escaping the baghouse were about 100 ppm, whereas proposed new requirements call for less than 30 ppm.

The Rotary Atomizing™ scrubber was installed to provide better control over emissions of acid gases, PICs, heavy metals, and particulates compared with the baghouse system. Installation and start-up of the skid-mounted EMCOTEK advanced scrubbing system at the commercial facility was accomplished by a three-man crew in less a week.

THE ROTARY ATOMIZING™ SCRUBBER

As shown in Figure 1, the Rotary Atomizing™ scrubber consists of a pre-spray tower, a Rotary Atomizer section, and a demisting section, followed by a fan and stack. The Rotary Atomizer creates a constantly renewed high velocity water curtain equivalent to a 5,000 psi spray nozzle. This water curtain is generated by means of a high speed rotating disc, as in a spray dryer. However, unlike a spray dryer, water droplets need to travel only a small distance before they hit the wall of the surrounding cylindrical tube, thereby creating a dense water curtain equivalent to a filter pad, which provides baghouse-equivalent particulate removal performance. Particulates passing through this concentrated spray curtain of atomized water droplets are removed by impaction, while gases are collected by absorption. Substantially greater particulate removal efficiency is provided, relative to power input, than high energy venturi scrubbers. In addition, the pressure drop through the Rotary Atomizing™ scrubber is generally about 4 inches of water, including demisting section pressure losses. Figure 2 presents a diagram of the Rotary Atomizer section. The water curtain can not plug or clog, unlike cloth or cartridge filters.

Following the atomizer, chevron-type demisters and droplet coagulating pads are used to collect and separate the water droplets from the gas stream. Single-stage demisters are normally rated for 95-98% removal per stage, of droplets 10 microns and larger. The scrubber in this application has two demisting stages, rated 99%+ overall.

In summary, the most significant advantages of the EMCOTEK Rotary Atomizing™ wet scrubber are the simultaneous removal of submicron particulate, condensible organics such as dioxin, and acid gases in a single integrated device, at a much lower temperature than a baghouse or ESP, and a far lower gas-side pressure drop than a venturi scrubber. It is forgiving and requires no operator attention as changes occur in the rate of burn, or nature of waste being incinerated.

ROTARY ATOMIZING™ SCRUBBER PERFORMANCE

Returning to Figure 1, one practical objective is disposal of the scrubbing water. At the demister, the scrubbing water must be as clean as possible, since any soluble salts carried over in the demister loss droplets are counted as particulate emissions. On the other hand, the volume of scrubbing water must be reduced to an absolute minimum. (At this facility, the bottom ash as well as the scrubbed fly ash are hauled to a Class I solid waste site). In order to meet the particulate removal and minimum liquid blowdown requirements, the EMCOTEK installation was designed with three circulating liquid loops, with the liquid being staged in a counter-current direction with respect to the gas flow. The first and cleanest liquid circulating loop consists of the Rotary Atomizers and the demister section. Fresh water enters this system on level control demand. Fresh caustic reagent is also introduced into this loop, based on pH control demand. Typically, the recirculating flow to the Rotary Atomizer is 14 gal/min, of which about 6 gal/min is fresh water.

Overflow from the first circulating stage enters the prespray tank, pumps, and spray tower, which constitute the second circulating stage. Washing the entering gases at an L/G ratio of 60 gal/macf, the nozzles in the spray tower knock down the mid-size particulate, and further neutralize the acid gases. The prespray outlet gases are completely saturated with moisture and are ready for "polishing" by the Rotary Atomizer. The prespray purge is a brine and ash solution containing 3 to 5% solids, which is fed to the quench system, at about 4 gal/min.

The dirtiest liquid is recirculated in the third liquid loop, consisting of an Inconel quench vessel, 4 feet in diameter and 14 feet high. During operation, the quench chamber serves three functions: (A) cooling the hot incinerator off-gases, (B) evaporating waste water, and (C) partial acid gas removal occurs in the quench vessel. Most of the hydrochloric acid and sulfur dioxide gases are converted to sodium chloride and sodium sulfate salts. Simultaneously, water is evaporated, allowing the soluble salts to build up to about 15% total dissolved solids (TDS) concentration. At 15% TDS, a blowdown of only 5 gal/hour, or less than 0.5% of the scrubber recirculating flow, was required.

This blowdown brine solution is blended with the bottom ash for disposal at a hazardous waste site. Alternately, it can be converted to a dry cake in a miniature, electrically heated evaporator. The total weight of the brine cake is just 6 to 7 pounds per hour. It represents only 10 to 12% of the weight of the incinerator bottom ash.

In order to minimize corrosion, the design includes appropriate materials of construction, redundant pH control, and the neutralization of acid gases as early as possible in the process. Once the gases have been quenched and neutralized, the relatively low temperature of scrubbing solution, 155 to 170°F., produces a less severe environment than many supposedly "dry" systems.

Test results comparing the baghouse system with the Rotary Atomizing™ scrubber system are compared in Table 1A and 1B. The Rotary Atomizing™ scrubber provides about 60% less total particulate emissions compared with the baghouse system. As shown in Table 1A, at reduced flue gas flowrates, the particulate removal performance of the Rotary Atomizing™ scrubber deteriorates, due to reduced gas velocities through the demisting section. The particulate material on the hot filter was determined to contain soluble salts (from the demister droplet losses), and these soluble salts were 76 wt% NaCl and 24 wt% Na_2SO_4. Even more surprising and important, the EMCOTEK Rotary Atomizing™ scrubber produced an extremely significant reduction in organic and condensible emissions, including dioxins and furans. As would be expected, the acid gas removal of the Rotary Atomizing™ scurbber was about 85% lower than the acid gas emissions of the lime-injected baghouse system.

PROCESS ECONOMICS

The baghouse contained 81 bags, 10 feet long and 6 inches in diameter, for a total bag area of 1260 ft^2. It was designed to operate handling a burn rate of 650 lb/hr, corresponding to 3600 ACFM at 400°F, with a pressure drop of 3 to 6 inches of water. Because of bag plugging, by January 1988 when using the baghouse, the incinerator was operated at only 300 to 350 pounds per hour. At this burn rate, about 2.5 pounds per hour of lime is pneumatically injected into the baghouse. At the same time, about 2 lbs per hour of caustic (100% basis) are used in the quench sprays to neutralize the acid gases (HCl, SO_2, et.) present in the flue gas. Power consumption for fans and pumps is 25HP. Bag replacement cost is shown on a per-ton basis. The afterburner (secondary chamber) consumes up to 4.4 SCFM of natural gas to maintain the incinerator exhaust temperature at 1800°F. minimum.

The wet scrubber operates at 3600 ACFM at 160°F. Its system pressure drop is 4 to 5 inches of water, and this airflow allows the incinerator to operate at 650 to 700 lbs per hour. Caustic consumption is stoichiometric, at 2.8 pounds per hour (100% basis). Power consumption for fan, rotary atomizer and recirculation pumps is 54 HP. Natural gas consumption of the afterburner is the same, per unit of waste, whether baghouse or scrubber is in use.

Table 2 presents an operating cost estimate comparison of the lime-fed baghouse and the ROTARY ATOMIZING™ Scrubber. At the same burn rate, costs are very nearly equal.

REGULATORY CLIMATE

Regulation of opacity and particulate emissions from incinerators has existed for decades. These regulations are now being gradually tightened worldwide, and other emissions, which in the past were simply worries or subjects of concern, are now being put under legal limitations.

In some parts of the industrialized world, identical rules apply for both municipal waste and biohazard incinerators. This is exemplified by the present U.S. EPA guideline of 0.08 gr/sdcf for all incinerators, soon to be lowered. However, in certain areas tighter rules apply to large MSW plants, than to hospital incinerators, because of the perceived advantages of incinerating biohazard material at or near the place of origin, rather than transporting it. Good examples of this are in the Los Angeles basin, and in Pennsylvania. In fact, the latest Pennsylvania rules embody a three-step approach, in order to avoid requiring high-cost pollution control at the smallest incinerators.

Table 3 shows typical regulation levels for MSW incinerators, and Table 4 for biohazard incinerators. These are compiled both from the literature (2,3,4) and from proposal requests EMCOTEK has received. An interesting pattern we note is that for biohazard incinerators, generally tighter limits exist in eastern USA and eastern Canada, than in the west. In fact, in California, there is still no statewide requirement for HCl control, except for a 4 lb/hr emission maximum imposed upon central collection facilities. Local districts regulating local hospitals have not had to require acid gas control, and this has often led to the installation of mild steel abatement devices intended to remove particulates only, which then fail by corrosion in short order. We expect this will change, as a 30 ppm HCl limit becomes uniform throughout the 50 individual California districts.

The tightest regulations for particulate emissions, in the range of 0.01 grain/dscf at 12% CO_2, in fact require unadjusted emissions of less than about 0.004 grain/dscf. With enough energy input, it is within the power of wet collectors to curtail acid gas emissions and at the same time attain these very low levels of particulates, as our data show. However, we must note that this places quite a burden on the designers of demisting equipment. Any droplets carried over to the stack contain salt at some low concentration, and any salt carry-over counts as particulate, even though all ash and soot emissions have been eliminated. Demisting efficiencies of 99% are no longer good enough. Increasingly, wet scrubbers have to be designed with two, three, and even more stages of demisting to eliminate all traces of salt-bearing liquid from the stack. From the standpoint of public acceptance, this is good, as ultra-thorough demisting and stack re-heat techniques eliminate the fall-out nuisance formerly associated with wet scrubbing.

We have seen specific, numerical limits on dioxin (TCDD equivalent) listed for Sweden and Pennsylvania, and on metals such as mercury and cadmium for Sweden, Germany, and Pennsylvania (2,3,5,6). In many other jurisdictions, these are handled on a case-by-case basis, as permits are issued by the authorities. Emissions of these toxic condensibles are substantially lowered by scrubbing at lower temperatures. For example, testing in Sweden of a pilot wet scrubber emitting saturated flue gas at 140°F showed total mercury emissions as low as 0.01 mg/Nm^3, whereas a Swedish facility with dry injection and a fabric filter operating at 320°F. emitted 0.065 mg/Nm^3 and in Germany a unit operating at 302°F. emitted 0.05 mg/Nm^3 or greater (2).

EMCOTEK believes that wet scrubbers operated at 140 to 160°F., will remove a much higher portion of dioxins, furans, and other chlorinated organics, because these compounds will have had a greater opportunity to condense on flyash particles or be absorbed by scrubber water at the lower operating temperature (7). Hence, the stack gas will contain less such pollutants than a dry system operated at 300-400°F. For example, a mass balance derived from the Hasselriis paper (8) indicates that up to 1500 ppt of such organics leave the stack with an ESP in the system. But when the vapor pressure is compared at 160°F, for typical dibenzofuran only 30 ppt should remain in the wet scrubber stack.

Wet scrubbers therefore can meet the tightening requirements of condensible toxics removal, better than dry systems. The EMCOTEK wet scrubber design can do so at low pressure drop.

SUMMARY AND CONCLUSIONS

A new type of wet scrubber providing baghouse-equivalent particulate removal has been developed and installed in a hospital infectious waste burning facility. This paper presents compliance test data obtained at this central-burn facility located in California. The test data includes both acid gas and particulate emissions based on operation of two separate flue gas cleaning systems. One system is a pulsejet baghouse equipped with pneumatic lime injection. The other system is the EMCOTEK Rotary Atomizing^{TM} scrubber which uses caustic reagent. Both systems were tested downstream from an Inconnel quench chamber.

At bio-hazardous waste burning rates which are nearly identical, the Rotary Atomizing^{TM} scrubber produced average particulate emissions of 0.022 gr/DSCF, corr. 7% O_2, approximately 60% less than the pulsejet baghouse system. The wet scrubber also produced 94% less organics and condensibles, such as dioxins and furans, compared with the baghouse system. The Rotary Atomizing^{TM} scrubber also produced 85% less acid gas emissions compared with the pneumatic lime-injected baghouse. The test results show that the wet scrubbing system using an EMCOTEK Rotary Atomizing^{TM} scrubber provides superior flue gas cleaning performance compared with a dry lime-injected baghouse system.

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The work described in this paper was not funded by the U.S. Environmental Protection Agency. The contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

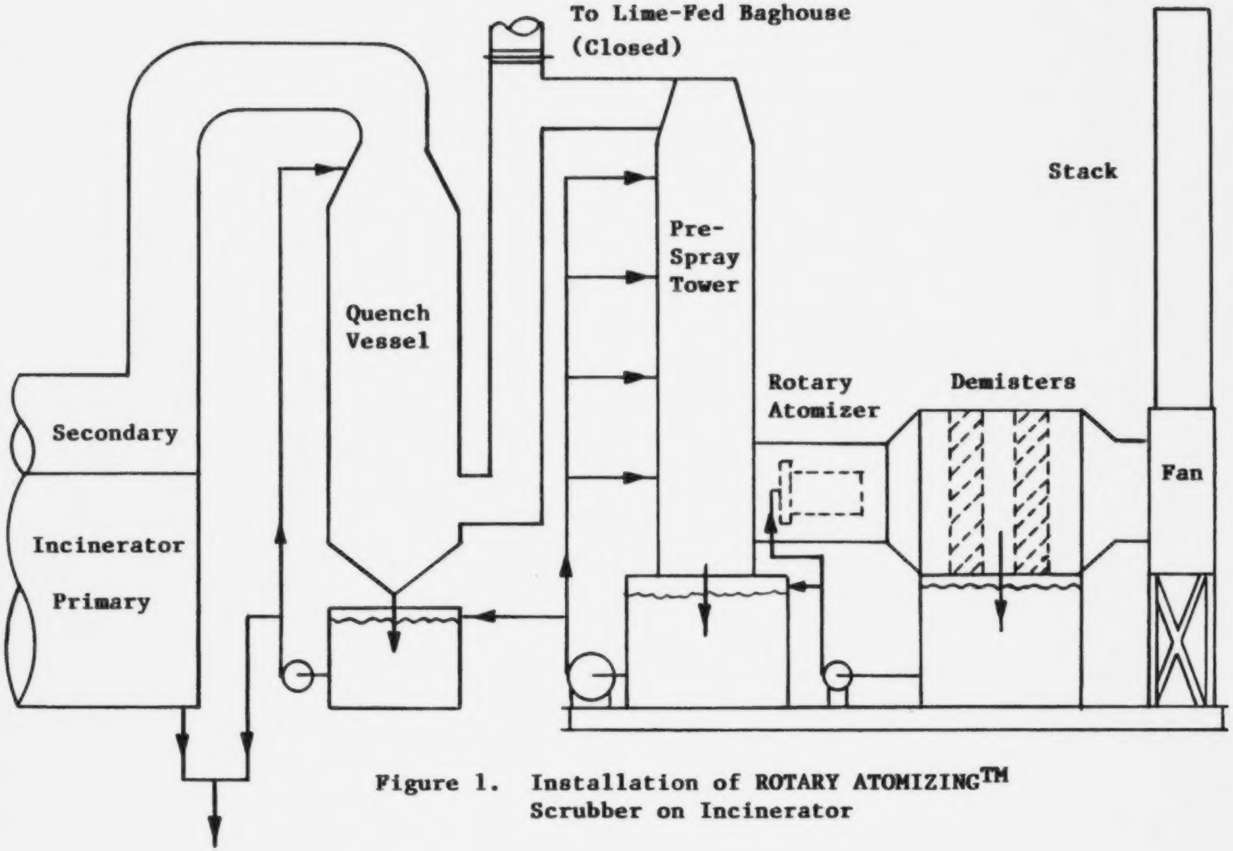

Figure 1. Installation of ROTARY ATOMIZING™ Scrubber on Incinerator

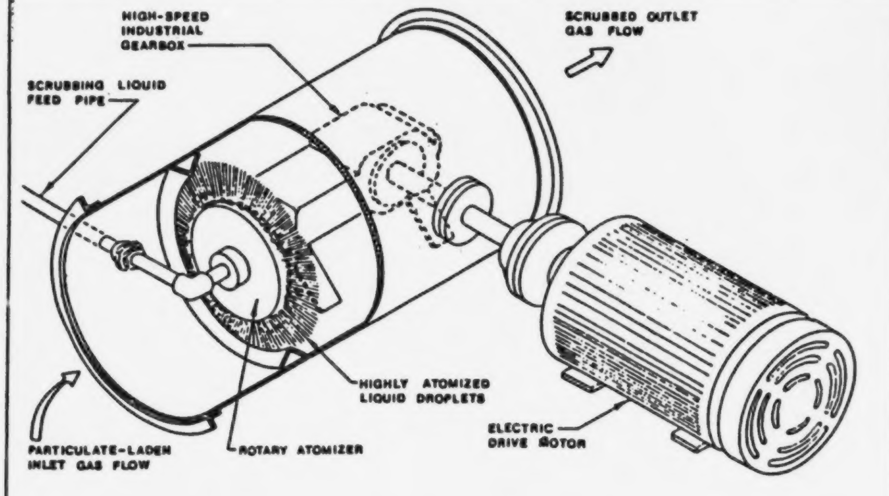

Figure 2. Internal Arrangement of the Rotary Atomizer Section

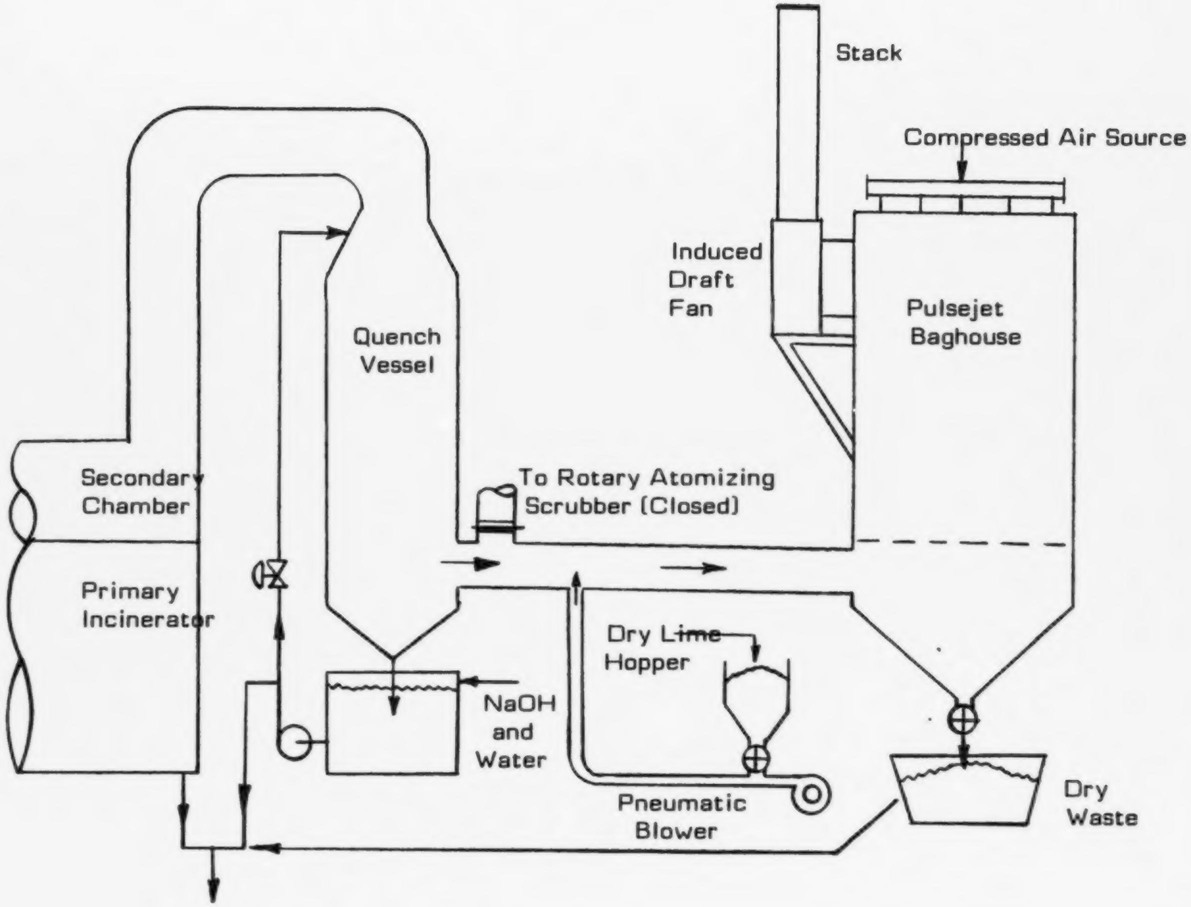

Figure 3: Installation of Baghouse and Lime Injection System on Incinerator

Table 1A: COMPARISON OF BIO-HAZARDOUS WASTE INCINERATOR EMISSIONS

	BAGHOUSE PERFORMANCE (Lime Injection)		ROTARY ATOMIZING^{TM} SCRUBBER PERFORMANCE (Caustic Reagent)		
Bio-Hazardous Waste Burning Rate, lb/hr	663	672	554	676	820
Stack Gas Flowrate and Composition					
ACFM	3400	3500	2900	4000	3700
Temperature, F	400	400	167	160	159
Moisture Content, vol.%	34.7%	34.1%	37.1%	37.3%	36.1%
DSCFM	1340	1400	1500	2100	2000
CO_2, vol.%, dry	4.8%	5.9%	6.8%	7.3%	5.4%
Oxygen, vol.%, dry	14.2%	13.2%	13.0%	11.5%	14.0%
Stack Particulate Emissions (Method 5), grains/DSCF					
Hot Filter, Insoluble	0.007	0.013	0.004	0.006	0.006
Hot Filter, Soluble Salts*	N/A	N/A	0.014	0.002	0.003
Organics & Condensibles**	0.010	0.025	0.001	0.001	0.001
Total Particulate, gr/DSCF	0.017	0.038	0.019	0.009	0.010
TOTAL PARTICULATE (corrected to 7% O_2)	0.035	0.068	0.033	0.013	0.020
Stack Acid-Gas Emissions					
HCl, ppm vol., dry	123	84	11	23	14
SO_2, ppm vol., dry	not measured		1	2	2

*Soluble salts come from demister losses and appear on the hot filter in the form of soluble salts. The analysis of these soluble salts was 76 wt% NaCl and 24 wt% Na_2SO_4.

**Organics and condensibles include dioxin, furan, hydrocarbons, fatty acids, phenols, etc.

Table 1B: COMPARISON OF BAGHOUSE VS. ROTARY ATOMIZING™ SCRUBBER PERFORMANCE IN A BIO-HAZARDOUS WASTE INCINERATOR APPLICATION

	Baghouse With Pneumatic Lime Injection	Rotary Atomizing Scrubber With Caustic Reagent	Rotary Atomizing Scrubber Improvement In Emissions Control Compared With Baghouse
Avg. Burn Rate	668 lb/hr	683 lb/hr	Burn Rates and Stack Gas Flowrates Are Approximately The Same For Both Types of Control System
Avg. Stack Gas Flowrate	3450 ACFM @ 400 F 1380 DSCFM @ 13.8% O_2	3530 ACFM @ 162 F 1860 DSCFM @ 12.8% O_2	
Avg. Stack Particulate Emissions:			
Hot Filter, Insoluble	0.010 gr/DSCF	0.005 gr/DSCF	50% less Insol. Particulate
Hot Filter, Sol. Salts	N/A	0.006 gr/DSCF	Increased Soluble Salts*
Organics & Conden.	0.018 gr/DSCF	0.001 gr/DSCF	94% less Organics/Condensibles**
Total Particulate	0.028 gr/DSCF	0.012 gr/DSCF	60% less Total Particulate
Total Particulate (corrected to 7% O_2)	0.052 gr/DSCF	0.022 gr/DSCF	
Avg. Acid Gases	HCl = 104 ppmv.dry SO_2 not measured	HCl = 16 ppmv.dry SO_2 = 2 ppmv.dry	85% less Acid Gases

* Soluble salts come from demister losses and appear on the hot filter in the form of soluble salts. The analysis of these soluble salts was 76 wt% NaCl and 24 wt% Na_2SO_4.

**Organics and condensibles include dioxin, furan, hydrocarbons, fatty acids, phenols, etc.

TABLE 2. OPERATING COST, AIR POLLUTION CONTROL, PER TON OF BIOHAZARD WASTE INCINERATED

	Lime-Fed Baghouse		ROTARY ATOMIZING™ Scrubber
Average Burn Rate, lb/hr	325	650	650
Electric Power	$6.87	$3.43	$7.42
Chemicals	3.06	3.06	1.91
Bag Replacement	3.44	3.44	-
Auxiliary Fuel Gas	10.40	5.20	5.20
PER TON, TOTAL	$23.11	$15.13	$14.53

At equal rates of incineration, operating costs are essentially equal.

TABLE 3. MAXIMUM EMISSION LIMITS FROM MUNICIPAL SOLID WASTE INCINERATORS, 1988

Jurisdiction	Particulates, gr/sdcf *	HCl, ppm
Sweden	0.008	63
West Germany	0.012	31
U.S.A. (EPA, current)	0.08	--
U.S.A. (NSPS, 1989)	0.046 §	--
California (ARB)	0.010 §	30 §
Pennsylvania	0.015	30
Connecticut	0.015	(-90%)
Michigan	0.015	(-90%)
California (L.A. Basin)	0.016	30 §
California (Mariposa)	0.02	--
California (other)	0.08	--
Oregon	0.02	--
Washington (Seattle)	0.05	50
Alaska	0.05	--
Canada (B.C.)	0.10	--

*Adjusted to either 12% CO_2 or 7% O_2, per jurisdiction.
§Proposed law, or guideline for permitting.

TABLE 4. MAXIMUM EMISSION LIMITS FROM HOSPITAL AND BIOHAZARD INCINERATORS, 1988

Jurisdiction	Particulates, gr/sdcf *	HCl, ppm
U.S.A. (EPA, Current)	0.08	--
U.S.A. (NSPS, 1989)	0.046 §	--
Pennsylvania:		
Under 500 lb/hr	0.08	(-90%)
500-2000 lb/hr	0.03	30
Over 2000 lb/hr	0.015	30
New York	0.015	50
Canada (Montreal)	0.022	50
Washington (Seattle)	0.03	50
Michigan	0.10	--
Texas	0.10	41
North Carolina	0.10	--
Canada (Alberta)	0.11	100
California (DOHS)	0.08	30 §
California (L.A. Basin)	0.10	30 §

*Adjusted to either 12% CO_2 or 7% O_2, per jurisdiction
§ Proposed law, or guideline for permitting.

THE DESIGN, START-UP AND OPERATION OF THE INTERMOUNTAIN POWER PROJECT, UNIT #2 FABRIC FILTER SYSTEM

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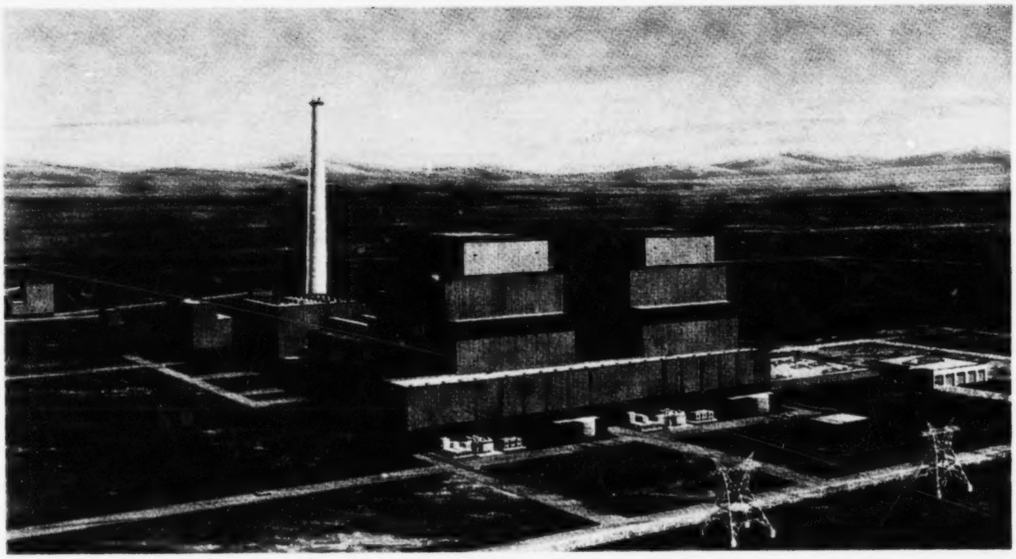

ABSTRACT

The Intermountain Power Project (IPP) consists of two (2) 840 gross MW coal fired boilers, located near Delta, Utah. This system represents one of the largest operating fabric filter systems in the world. Unit #2 has been successfully in operation since February 16, 1987. Special procedures implemented during the design phase, and detailed quality assurance practices during fabrication, which in large part are responsible for the success of this installation are reviewed. Unit #2 design, start-up and operating experience, including test results, will be presented in detail.

INTRODUCTION

The Intermountain Generating Station (IGS) consists of two (2) 840 MW gross (800 MW net) pulverized coal fueled steam generators designed to burn Utah coal. Each generator has a maximum continuous rating (MCR) of 6,100,000 lb/hr steam at 2,640 psig with a gross heat input of 8,352 million Btu/hr. While both units were originally designed at 820 MW gross rating, they have since been upgraded to the existing rating. Each unit is equipped with an air quality control system (AQCS) consisting of a fabric filter (baghouse) for particulate control, a wet scrubber for removal of sulfur dioxide, and a sludge conditioning system for disposal of combustion wastes.

Initial flue gas operation of the Unit #1 baghouse (4) began on March 27, 1986 with commercial operation in June, 1986. Unit #2 baghouse began initial operation on February 16, 1987 with commercial operation on June 13, 1987.

This station is located 100 miles southwest of Salt Lake City, near Delta, Utah. The station owner is the Intermountain Power Agency (IPA), which is a subdivision of the State of Utah. Project management is provided by the Los Angeles Department of Water & Power (LADWP), who is also the largest single purchaser of Intermountain generating capacity. The architect engineer was Black & Veatch (B&V).

Particulate emission guarantees are as follows: The total emission rate at the stack shall not exceed 0.020 lb per million Btu of heat input to the steam generator, and opacity of the flue gas at the stack shall be no more than 20 percent opacity. Results of the formal contract compliance testing conducted during July, 1987, are presented in Table 4.

GE Environmental Systems provided each of Units #1 and #2 with a single baghouse system consisting of three (3) separate casings, with sixteen (16) compartments per casing, each containing 396 filter bags, for a total of 6,336 bags per casing or 19,008 bags per boiler. System design parameters are presented in Table 1.

SCOPE OF SUPPLY

The scope of supply for the baghouse system was from the inlet flange to the outlet flange of the individual casings including all structural steel, insulation, hoppers, individual compartment inlet and outlet manifolds, bypass ducts, poppets, reverse gas fans and motors, compartment ventilation systems, model study, filter bags, hopper heaters, and control enclosures with all devices for automatic and manual control, including state-of-the-art touch screen computer controls.

MODEL STUDY

A 1:16 scale model study was performed to simulate the gas and dust flow performance. The model study consisted of two (2) phases. The first phase consisted of tests on the flange-to-flange fabric filter casing model under various conditions of operation. These investigations were conducted to determine the gas flow division between the individual casing compartments and to make any physical changes required to obtain satisfactory balance for the various conditions. For the various operating modes investigated, gas flow division between compartments, pressure loss and flow/dust deposition were measured.

The second phase of the model study investigated the ductwork external to the flange-to-flange baghouse. This was conducted to optimize the configuration of the ductwork system, in regard to controlled gas flow. This included gas flow distribution within the ductwork, incremental ductwork system pressure drop characteristics, and biasing of gas flow rate to individual casing and I.D. fan components. These tests were conducted from 50% to 100% MCR system operation. Photographs of the models used in our study are shown in Figures 1 and 2.

BAG DESIGN

The filter bags used are twelve (12) inches in diameter by thirty three (33) feet long. The filtration material is 13.5 oz/yd sq. nominal weight woven fiberglass with a Burlington I-625 G acid resistant finish. The filter bags were manufactured by Midwesco Filter Media Resources Division in Winchester, Virginia. Each bag was supplied with eight (8) anti collapse rings located at variable spacings along the length of the bag, and is attached to the thimble without the use of any special tools and/or clamps. To ensure meeting the highest filter bag standards, a stringent bag/fabric QA/QC program (1) was conducted by an independent third party test laboratory. Verification of all tests results required were conducted by (LADWP) and (GE Environmental) QA/QC personnel. A summary of the fabric testing acceptance criteria is presented in Table 2.

Filter Bag Quality Assurance/Quality Control Program

Quality control procedures were performed by the manufacturers of the fabric, sewing thread, caps, rings, and completed filter bags. The quality assurance was performed by LADWP, GE Environmental, and an independent test laboratory approved by all parties. All the filter bags were manufactured to meet or exceed the requirements of a detailed and exhaustive specification prepared by GE Environmental and approved by LADWP and Black and Veatch (B&V).

Initially it was agreed that the test criteria was to be compared to the average of the required three test results per sample. However, multiple point averaging can mask the effect of one failing test point. Failures can occur at defect areas which could compromise proper filtration ability. To minimize that possibility, the test criteria was modified to take the form of: The average of the three test results must be at least (X) and each individual test result must be at least 90 percent of (X). After this modification was made, LADWP and GE quality assurance compared each test result as well as the average of the three test results to a specific test criteria for acceptance or rejection of goods. This allowed most of the material and bags to pass a very stringent test program but did not allow a low test result to be averaged out and accepted.

As a result, the QA/QC group achieved its goal of producing bags of consistent and high quality. The actual bag performance and longevity, however, are determined in the field and are affected by parameters such as the type of coal being burned and plant operation, which are not easily correlated to laboratory tests.

Bag Installation

The filter bags on Unit #2 were installed approximately two (2) months prior to start-up. Procedures required immediate sealing of the compartments after the bags were tensioned. Details of the bag hanging procedure are as follows:

> **STEP 1** - Confirm that the hopper heaters are operable in the compartments being bagged. Any thimbles with a bent lip or rough spots should be repaired and ground smooth prior to installing the bags to avoid damage to the bag upon installation.
>
> **STEP 2** - The boxes of the filter bags are to be opened and the bags laid out still folded on top of the flattened boxes as shown in Figure 3. This will prevent abrasion of the glass filter bags while resting on the thimbles. The bags laid out along the outer walls of the compartments are placed about two (2) feet in from the walls so that they do not catch on the insulation support pins nor horizontal support members while the bags are being raised into position and hung on the bag supports.
>
> **STEP 3** - As demonstrated in Figure 4, the filter bags (3) are pulled up, three at a time and attached at the top on the bag support bars, with the bag vertical seams facing the walkways. As soon as a quantity of bags have been pulled up and attached, the bag bottoms are placed on the thimbles as follows:
>
> - The bag cuffs are placed over the thimble lip at an angle sufficient enough to allow it to be installed.
> - Once the bag cuff is below the lip and wide part of the thimble, pull vertically upwards on the bag to secure it on the seat of the thimble. Make sure to check the full 360 degrees of the bag so that the bag compression band is up against the lip of the thimble. Check to ensure that the vertical bag seam is straight and not twisted.
>
> **STEP 4** - After the compartment has had the bags installed, the bags are fully tensioned to its preset tension of 75 pounds with the use of a pneumatic tensioning tool. The bags are tensioned from the outside walls in towards the center of the walkways to allow for any take-up in the tube sheet.
>
> **STEP 5** - After all the bags in the compartment have been installed and tensioned, the compartment is cleaned of any debris and sealed. The compartment doors are then tack welded shut and the hopper heaters turned on until after start-up to prevent unauthorized entry into the compartments and to prevent moisture condensation from occurring on the bags.

SYSTEM CONTROLS

The IPP control system consists of conventional control panels with hand switches, lights, indicators, recorders, annunciators, etc. and a "state of the art", cathode ray tube (CRT) based graphic system.

The "touch screen" graphic system duplicates both control and indication, as well as the annunciation provided by the conventional control panels. One graphic system with redundant CRT units was provided for the baghouse system. The graphic systems communicate via a programmable logic controller (PLC) located in each of the conventional control panels. Either the graphic system or the conventional control panel can operate the baghouse independent of each other. An example of one of the baghouse CRT screens is shown in Figure 5.

The touch screen graphic system allows the operator to control, to observe the system status and alarming of key components of the baghouse. The operator can within seconds, move from one component of the system to another by just touching the screen at the appropriate location. The requested control function can also be performed by selecting the required screen and component and then touching the graphic symbol on the screen. The function requested is then performed by the PLC. Alarms are displayed and acknowledged on all graphic screens as well as the main alarm screens. Active messages, as well as acknowledged and reset alarm messages, are displayed showing the time, day, message, and graphic number where the respective system is displayed.

CLEANING PHILOSOPHY

At IPP, the three (3) individual casings are treated as a single baghouse system. When any casing reaches its preset pressure drop initiation level, all three (3) casings are cleaned on a staggered basis, starting with casing "A". This patented (1) cleaning method staggers the casings 70 seconds apart from each other, thus allowing the full flow of reverse gas into only one (1) casing at a time. By doing this, the maximum system pressure spikes that accompany a cleaning are minimized. The cleaning cycle can be initiated by either 1) Casing Differential Pressure; 2) Preset Timer Override; or 3) Operator Initiated (Manual) Cleaning.

As demonstrated in Figure 6, Casing B will start cleaning 70 seconds after Casing A begins cleaning, and Casing C will start cleaning 70 seconds after Casing B has begun cleaning. The cleaning cycle time per compartment is 200 seconds as shown in Figure 7, with the compartments being programmed to clean from front to back as follows: 1,9,2,10,3,11,4,12,5,13,6,14,7,15,8,16. At this point, the baghouse will revert back to the filtering mode (All Compartments On-Line) until it again has its cleaning cycle reinitiated.

OPERATING HISTORY

PRECOAT

The filter bags were precoated with fly ash just prior to initial start-up. The purpose of the precoat was to protect the clean surface of the new filter media from any potential harmful constituents of the flue gas and/or boiler upsets. The filter bags were estimated to require approximately 1.5 pounds of dry fly ash per 10 sq. ft. of cloth. With a total of 1,970,608 square feet of cloth per boiler, about 1,478 tons of fly ash was required to effectively precoat the filter bags to a required pressure drop rise of 0.5 to 2.0 IN. W.C. over the clean bag pressure drops. The precoat material for Unit #2 was supplied from the Unit #1 baghouse hopper ash silos. The particle size distribution for this ash is shown in Figure 8.

All hopper heaters had been on since the bags were initially installed and the compartments sealed. This ensured a hot, dry environment in the baghouse compartments to prevent moisture condensation on the fabric. Air temperatures above 140 degrees F. were observed in the compartments.

Precoat began on February 12, 1987 at 9:54 AM and was completed by 4:00 PM that same afternoon. Figures 9, 10, and 11 show the initial clean bag pressure drops (Solid Bars) vs the pressure drops just after precoating (Hatched Bars) taken at design compartment flow rates. A total of four (4) compartments were precoated in each casing at a time to ensure obtaining the required flow rates. Prior to injecting the fly ash into test ports located in the inlet manifolds, the boiler was operated on oil, with the baghouse on bypass until gas temperatures were in excess of 200 degrees F.. The boiler flames were then extinguished and clean, hot air was pulled through the boiler into the baghouse to transport the ash (precoat material) into the compartments during precoat. If the inlet temperature dropped below 150 degrees during this process, the precoat was stopped and the boiler heat up process repeated until all the filter bags were precoated.

INITIAL OPERATION

The Unit #2 baghouse system was officially placed into operation on February 16, 1987 at 1:40 PM. The generator load was 400 MW, with casing pressure drops at approximately 1.7 inches W.C. and inlet temperatures ranging from 225 to 260 degrees F.. Generator loads remained constant for about 2 days and then were increased to 600 MW on February 18th. The first pressure initiated cleaning cycle took place at 7:30 AM on the 18th., with the baghouse system pressure drops reaching the required preset initiation level of 5.4 inches W.C..

The first compartment to see flue gas was compartment 16 in "B" casing. This compartment had no filter bags installed initially as it was used as a condensation vent to allow a free path for the initial quantity of flue gas to pass through and carry with it any condensation that may have occurred in the inlet manifold prior to startup. This compartment was later installed with filter bags following a required 500 hour bag seasoning period.

COAL SUPPLY

The initial coal supply at the IPP facility is mined in the State of Utah. This coal is a low sulfur, high grade, bituminous coal from underground mines located about 200 miles east of the site. Approximately 4.4 million tons will be required each year which equals 12,000 tons of coal daily. A typical fuel analysis for this coal is shown in Table 3.

PRESSURE DROPS

The system pressure drops have been well within the contract requirements since initial operation. Boiler loads have continuously increased on the average since startup. Figure 12, shows the average monthly peak pressure drops, megawatts, opacities, and number of compartments off line for maintenance vs time since startup in February of 1987. Both boiler loads and pressure drops can be seen to have leveled off since November while at the same time opacities have decreased to less than one (1) percent as a normal reading.

PERFORMANCE TEST RESULTS

Table 4, represents the formal contract performance test results conducted in July, 1987. Test results indicated that the baghouse system is operating extremely well, with time average pressure drops measured during the test period over two (2) inches below guarantee levels and particulate emissions measured at 0.005 lbs/MM Btu's exiting the baghouse. Opacities during the test ranged from 1.5 to 3.0 percent at the stack exit.

BAGHOUSE MAINTENANCE REQUIREMENTS

THIMBLE REPAIRS

Since startup, a small percentage of bags have come off their respective thimbles causing other adjacent bags to fail as a result. This was caused by a small percentage of thimbles being out of the specified tolerance range of 11.825 to 11.855 inches diameter. Even though the degree of variance from this specification was extremely small, with a clampless thimble/filter bag design, this dimension is critical.

This problem was initially discovered in Unit 1 and as a result, all thimbles were measured with a pie gauge tape prior to installing the filter bags on Unit #2. Out of 19,008 total filter bags, only 22 have actually come off the thimbles. These thimbles were repaired in place with a thimble enlarging device as shown in Figure 13. This device was designed and built by GE field construction, to actually cold roll the thimble back to within required tolerances. This device consists of rollers which can be adjusted to push out equally on three (3) sides of the thimble. By turning this device within the thimble with the use of a ratchet, the entire thimble can be expanded quickly and uniformly. These thimbles were repaired and no bags have since come off. As a result of this problem, a total of 108 bags have been replaced over the one (1) year period since startup. It should be noted that most of these bags did not actually fail due to their own defects, but were damaged by the action of a failed bag nearby. While these bags should never have been damaged in the first place, this equals only 0.6% of the total number of bags replaced to date.

HOPPER GASKETS

The original hopper door gaskets were to have consisted of a 3/4" closed cell Viton Sponge gasket. This was replaced with a fiberglass braided gasket due to problems discovered on Unit 1, where high temperatures in the hopper area surrounding the hopper access doors caused the Viton material to get hard and crack. This high temperature was caused by the improper location of the hopper heater temperature sensing probe which allowed the hopper heaters to remain on continuously. The probe is located in the hopper throat area where the least amount of heat is present. These heaters provide 40 KW per hopper which can generate localized skin temperatures in excess of 400 degrees F. Prolonged exposure of the gasket material to this temperature exceeded the design capability of this material. Based upon this experience, Unit #2 hopper access doors were installed with the fiberglass material during the final construction phase prior to startup. No problems have been encountered since this change was made.

REVERSE GAS DUCT EXPANSION JOINTS

Once again, based upon operating experience from Unit #1, the originally supplied Butyl rubber reverse gas expansion joints were replaced with a Viton reinforced material in anticipation of higher than designed operating temperatures in this area. The original joints failed due to cracking caused by temperatures exceeding the rated temperature of 325 degrees F. which was the original bypass high temperature set point. No problems have been encountered on Unit #2 with this new material as of this date since startup.

COMPARTMENT PURGE VENTILATION SYSTEM

To prevent potential vibration problems in the purge ventilation ductwork system caused by air turbulence, Unit #2's ventilation system was retrofitted with turning vanes and perforated plates to streamline the flows exiting the compartments into the roof mounted purge/ventilation ductwork, prior to exiting the vane axial fans. This vibration may have contributed to failure in the fan on one (1) casing on Unit #1. The turbulence was discovered as a result of a model study of the purge ventilation system conducted at the GE facilities. These modifications have resulted in low vibration levels at the axial fans which eliminate any potential fan failures in the future. To date, no failures have occurred on Unit #2. This action was of a preventative nature.

HOPPER ASH EVACUATION SYSTEM

The hopper ash evacuation system continues to be the cause of most of the compartment down times. Both Units #1 & #2 have been continuously plagued with high ash levels due to ash feeder gate problems causing them to jam shut. These valves have been rebuilt.

Another area of failure in the hopper, has been in the pressure equalizer lines. These lines shaped like an upside down U, developed holes in the upper knee of the curve within the hoppers, which resulted in ash jetting up towards the bags, through the thimbles resulting in five (5) bag failures to date. These problems have been minimal on Unit #2 as modifications have been made based upon Unit #1 experience.

AVAILABILITY IMPROVEMENT TASK FORCE (AIP)

The Intermountain Power Project is a prime example of what a good working relationship between customer, architectural engineer, and equipment supplier can achieve. Each organization can contribute to produce a final baghouse design and operation based upon industry and personal experiences.

Problems are always preventable in hindsight. However, given the importance of achieving reliable operation, a new approach to problem prevention was implemented. This was done through the formation of the Availability Improvement Task Force (AIP), which involved joint representation from LADWP, B&V, and GE Environmental along with outside consultants when deemed appropriate. Regular meetings provided a forum for ideas, changes, and/or concerns to be expressed and actions to be identified based upon facts vs speculation. Task items were identified, logged and the parties assigned the task of accumulating the appropriate data to either defend the proposal, reject it, or pursue additional information to achieve the best decision.

This approach worked extremely well on this project. Some of the areas investigated included hopper heater sizing; filter bag material including weight, construction, and finish; and even the broken bag detection system. These are just a few of the vast number of topics that were investigated and implemented by this committee. Although many were not chosen for implementation, the disciplined evaluation ensured that each item was given equal weight in the final consideration.

SUMMARY

Since startup, the Unit #2 baghouse system has successfully been in operation with few major problems. The unit has operated in compliance well within contract requirements of 0.020 lbs/MM Btu. and less than 20% opacity. System average pressure drops were measured over two (2) inches below the guarantee level of 6.8 inches W.C., and bag failures have been less than 0.6% of the total number of bags installed since startup. Based upon performance to date, it is anticipated that this system will continue to operate well.

It is likely that the successful operation of Unit #2, is the result of many factors beyond the design, including the coal being burned and the startup/operation procedures. All of these items contribute to a successful installation.

One major difference between both Unit #1 and #2 that may have contributed to the lower filter cake pressure drops and reduced bag failures is the actual bag finish.

A change was made in the supplier of the graphite component of the original acid resistant finish (GE I-625) which was supplied on the Unit #1 fabric. This change occurred between the time that the two fabric runs were being made at the weaver. What resulted, was a dramatic change in the fabric strengths and the low end of the permeabilities measured during the QA/QC program for the Unit #2 fabric/bags. This abrupt change from what we measured on the Unit #1 fabric, is shown in Table 5.

This table shows the low and high average values for the major tests conducted during this program for both the Unit #1 (GE I-625) and Unit #2 (I-625 G) fabric finishes.

While we can only speculate on whether these changes are directly responsible for these benefits, we believe that they may have at least contributed to these better results for the following reasons:

1) The low end of the permeabilities was raised from a minimum of 36.2 to 41.8 cfm/sq. ft., which may provide a lower clean bag pressure drop and higher flow throughput.

2) The new graphite, may have provided a greater surface area for the Teflon component to attract and adhere to, thus providing better dust release characteristics.

3) SEM photographs (6) taken of the two (2) finishes, showed a much smoother finish on the fibers with the (GE I-625), verses the (I-625 G) finish which reduced the amount of exposed textured fibers.

Of course, there are other factors that could have provided the same results, but, this is one change that we cannot overlook, and should be the source of further discussion.

To summarize, we believe that the key items that have contributed to the success of this project are as follows:

Proper System Design: The design of the system should be a joint effort based upon the accumulated experiences of the industry as a whole, including the specific experiences of the equipment supplier, the architect-engineer, and the customer. All have relevant experience covering different perspectives which, when combined, yields the best possible design.

Availability Improvement Task Force (AIP): This committee helped ensure that the best of all technologies was considered in the final design. This included controls, materials of construction, bag materials and coatings, reviewing the final results of the bag QA/QC program team, startup procedures, and general operation.

Detailed System Model Study: The model study has again proven itself to be valuable in proper baghouse system design. This provides verification of mechanical pressure drop and flow balances, prevents excessive drop out of fly ash in the ductwork, and improves overall mechanical system pressure loss through optimized vaning and distribution devices.

Proper Bag Design: The final selection of the fabric material, weight, weave, and applied finish has a great impact on the general operation of the baghouse and its tolerance to upset conditions such as boiler trips, tube leaks, and cleaning forces. The use of the heavy weight (13.5 oz./sq. yd.) fabric may have contributed to the extremely low particulate emission levels and the (I-625 G) acid resistant finish to its tolerance to boiler upsets and possibly even its low bag pressure drops.

Bag QA/QC Program: This program ensured materials were within design specification. The program, as described in this paper, covers all aspects of the bag design including the finish, weave, material strengths, weights, caps, threads, rings, and bag-to-thimble fit. This fit is essential in ensuring a tight seal when utilizing a clampless thimble design.

Proper Bag Installation: The bags must be installed properly on the thimbles and tensioned to their required force. Without proper tension, the bags could become loose during operation and actually come off the thimble causing surrounding filter bags to be damaged. The bag-to-thimble fit is critical in obtaining low emission levels as demonstrated on Unit #2. Final results are indicative of 100% of all 19,008 bags having an acceptable fit and proper seal. Bags were retensioned if required after the bag seasoning period was completed.

Ideal Coal For Baghouses: A coal low in moisture, sulfur, sodium, and relatively high in calcium and magnesium, may compliment the ability of a baghouse to provide low pressure drops, low emission levels, and low bag weights.

Precoat Requirement: There are many arguments both pro and con relative to the use of precoat requirements. It can be argued that it is better, not to precoat, but to bring the baghouse on line with its own boilers fly ash. Others sources (5) say that you should protect the fabric as much as possible using a known source of material having the ideal ash properties. For the IPP installation, it was decided to precoat the filter bags using the best ash possible. While on Unit #1, the ash required was transported in from another boiler installation, on Unit #2, we used the fly ash from right from the Unit #1 ash storage silos. We believe that the use of precoat has provided protection during startups and shutdown from the condensation of acids and/or moisture. This has also provided protection during boiler upset conditions such as blackouts when purging of the baghouse wasn't possible.

Proper Startup Procedures: Proper startup procedures contributed to the successful operation of this baghouse system. These procedures included: 1) Preheating the casings above the moisture dew point before exposing the bags to flue gas, and 2) Venting the initial volume of flue gas that is collected in the inlet manifold through one specific unbagged compartment. (At IPP, compartment B-16 was used as a condensation vent).

These are key factors in a successful baghouse installation. A lot of thought went into the quality assurance of all these items since the beginning of the project. They, in combination with the right amount of serendipity, have yielded a superb operating system.

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3. Figures 3,4, and 5 Courtesy of ICI Advanced Materials. Suppliers of FLUON PTFE To Burlington Glass Fabrics.

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The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

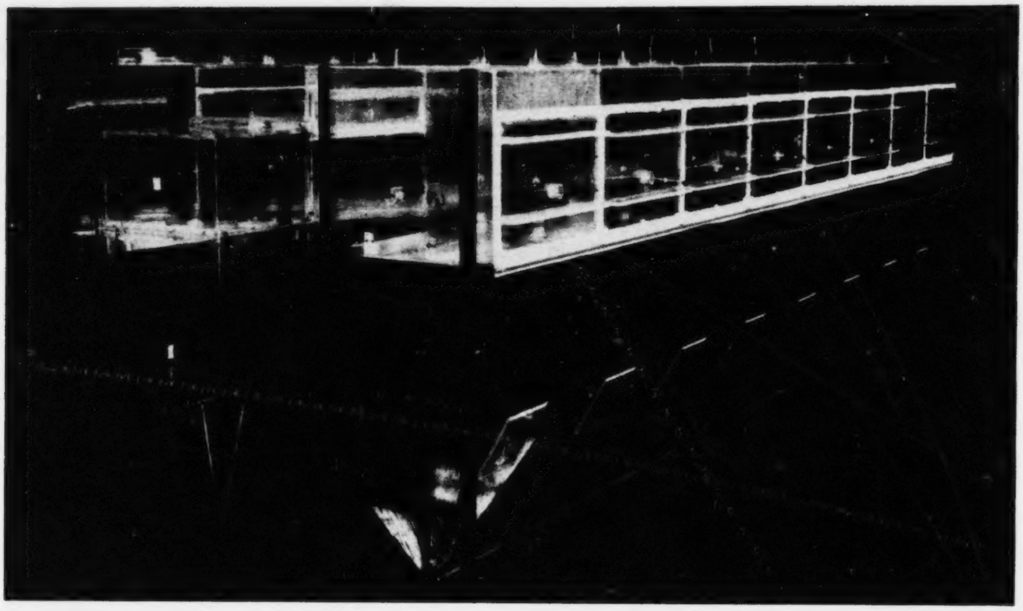


Figure 1. Flange to flange baghouse casing model.

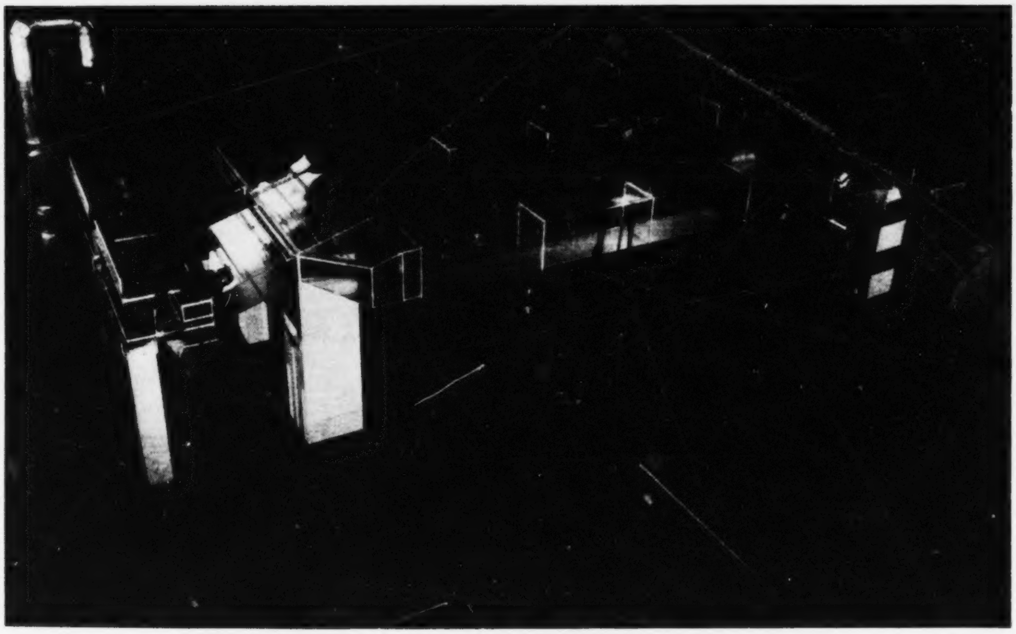


Figure 2. Total baghouse system ductwork model.

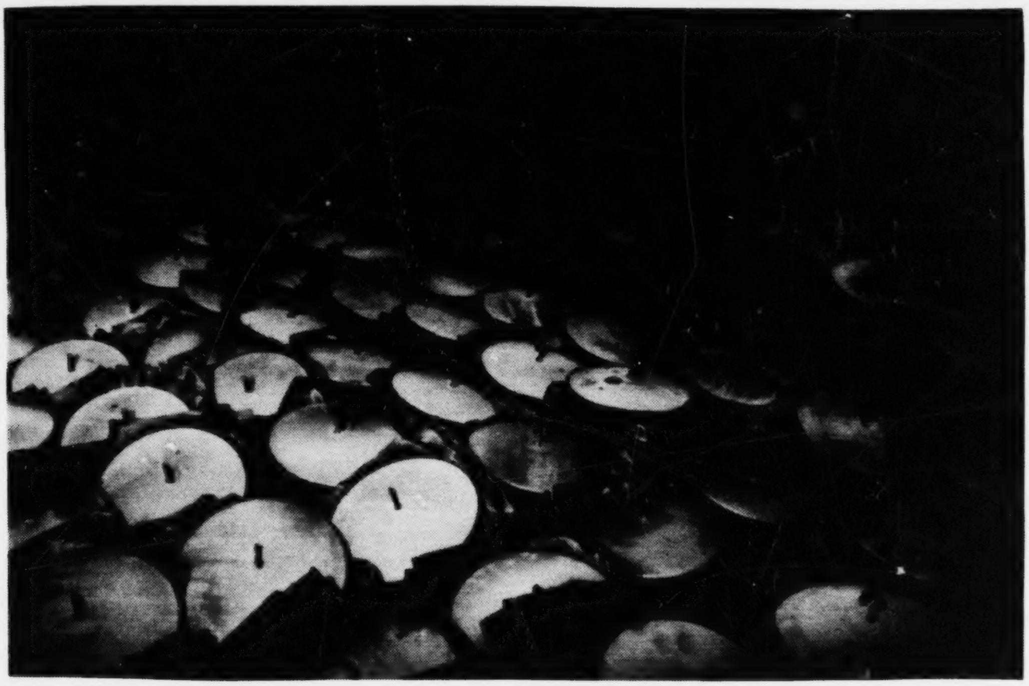


Figure 3. Filter bags laid out prior to hanging .

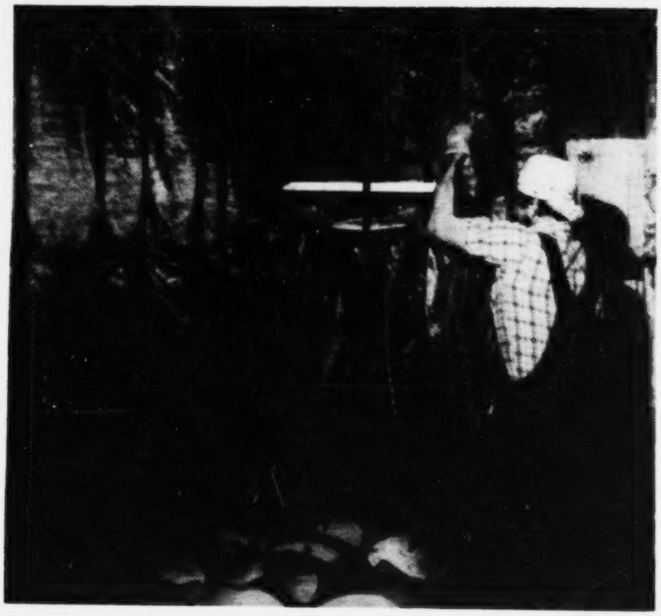


Figure 4. Filter bags being raised to upper support level.

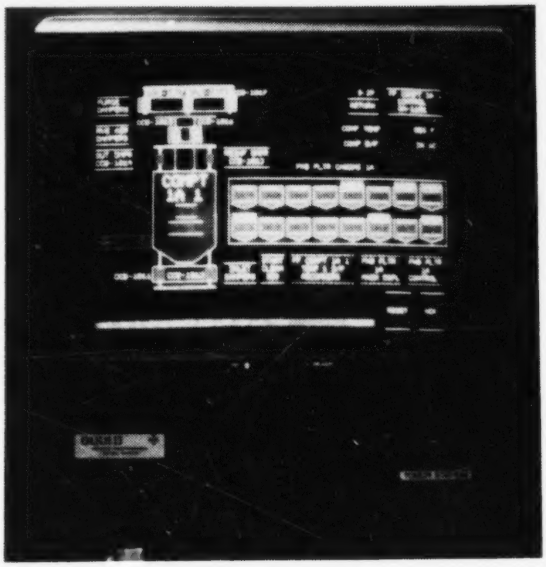


Figure 5. Touch screen CRT control monitor.

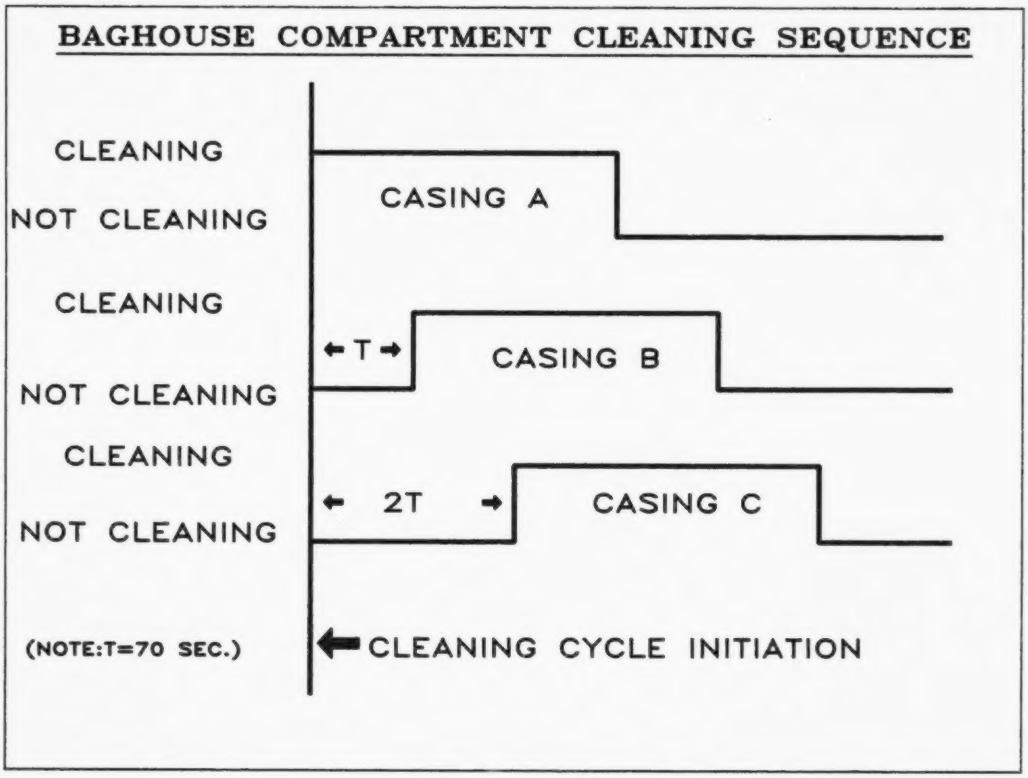


Figure 6. Baghouse system compartment cleaning sequence.

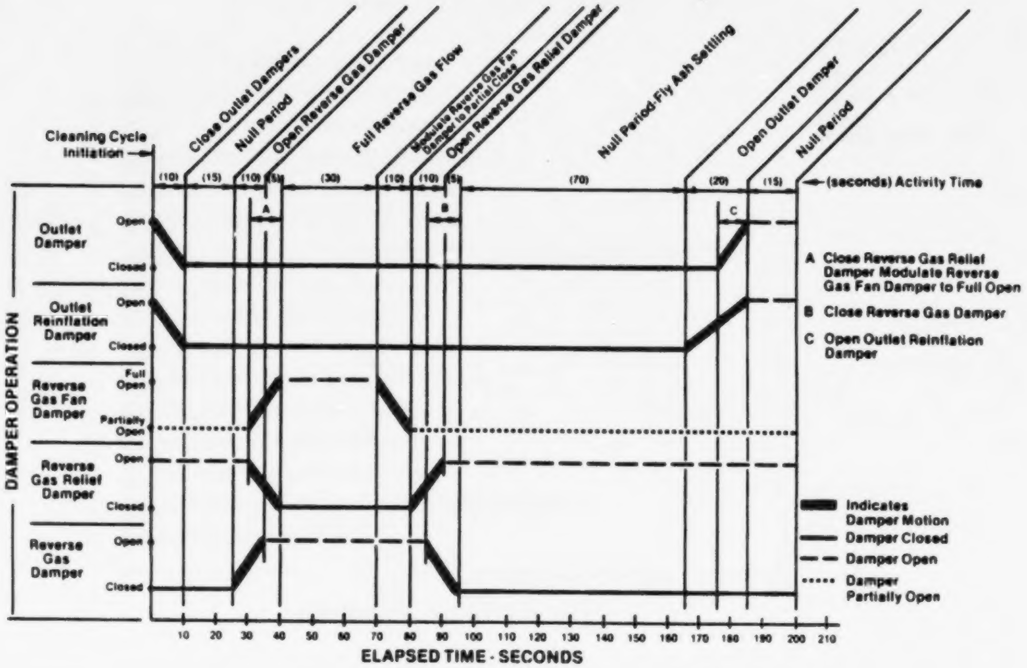


Figure 7. Individual compartment cleaning sequence,

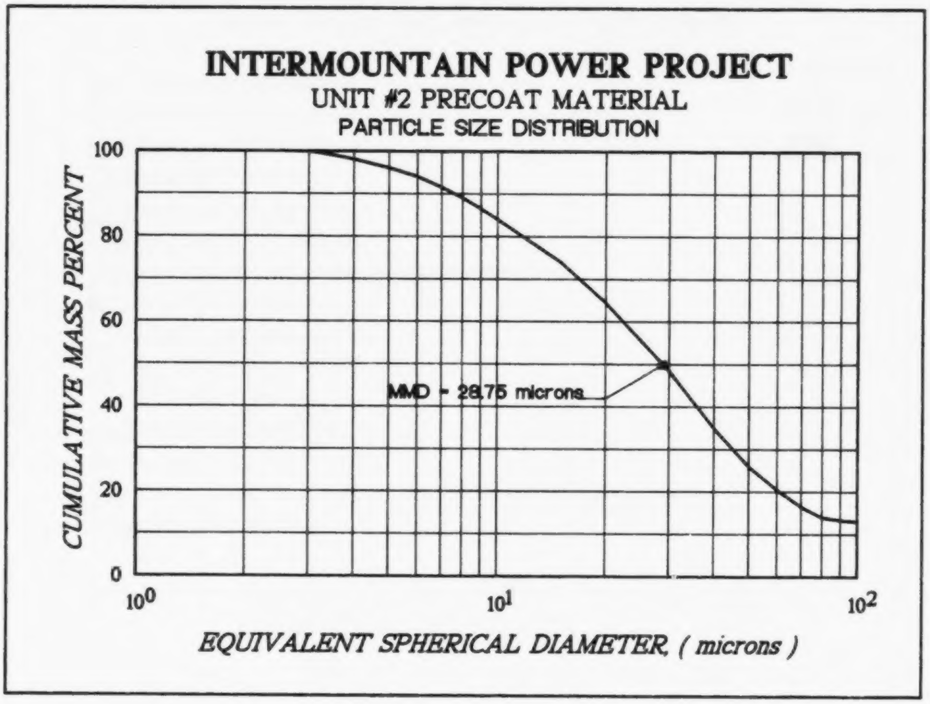


Figure 8. Unit #2 precoat material particle size analysis.

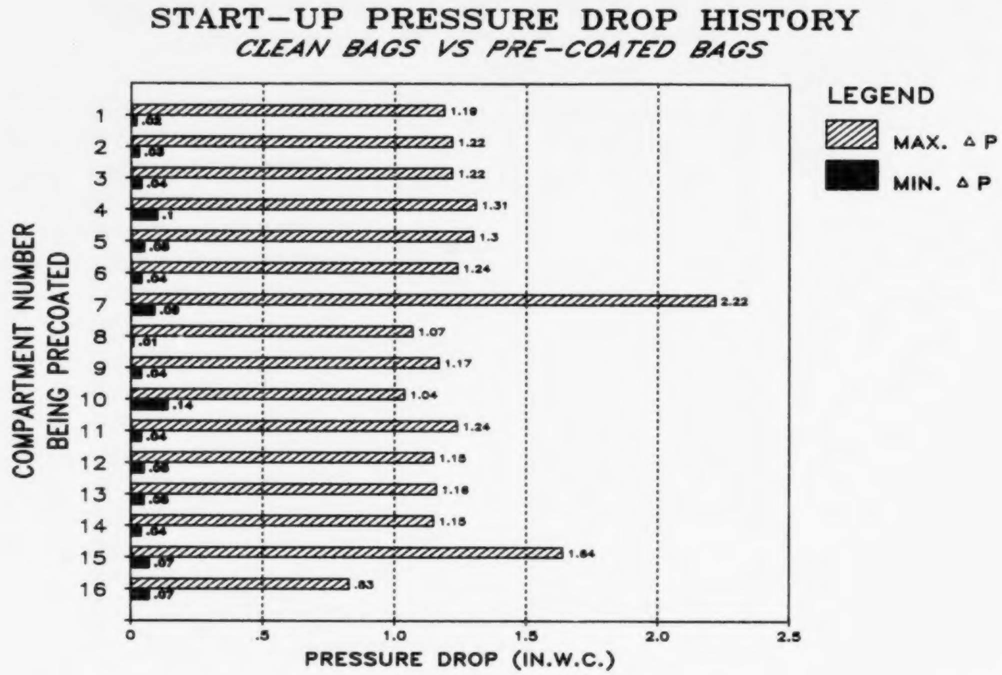


Figure 9. Casing 2A startup pressure drop history.

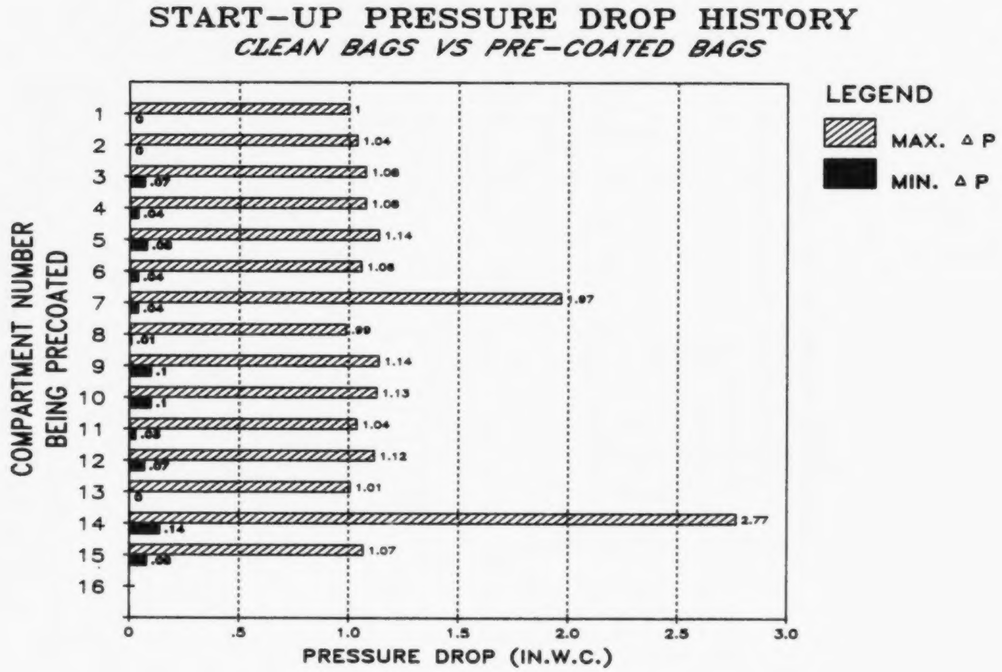


Figure 10. Casing 2B startup pressure drop history.

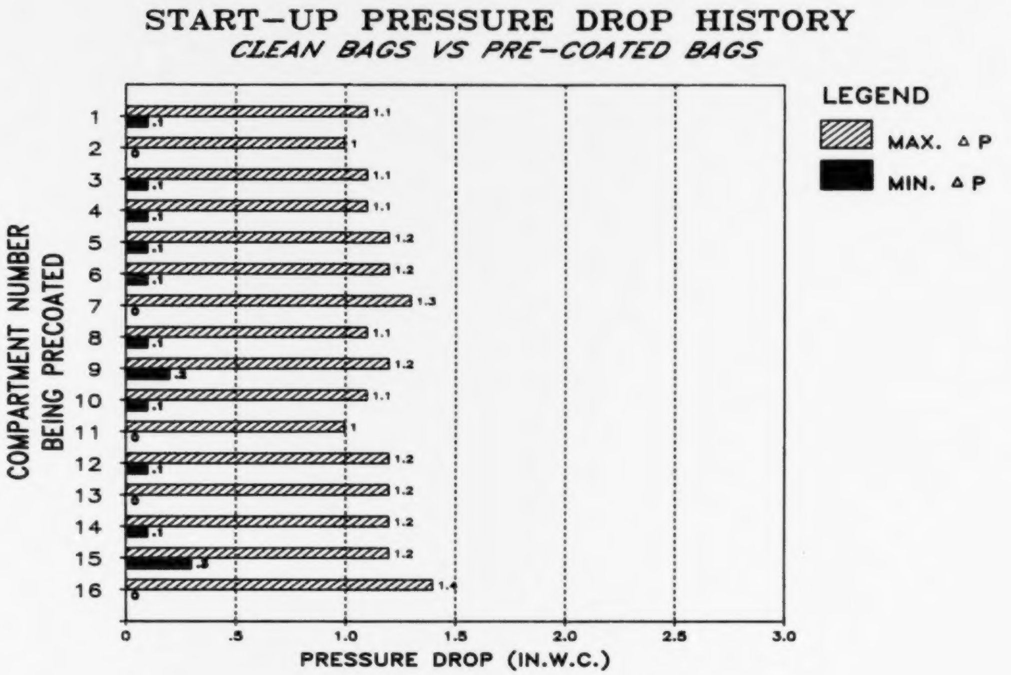


Figure 11. Casing 2C startup pressure drop history.

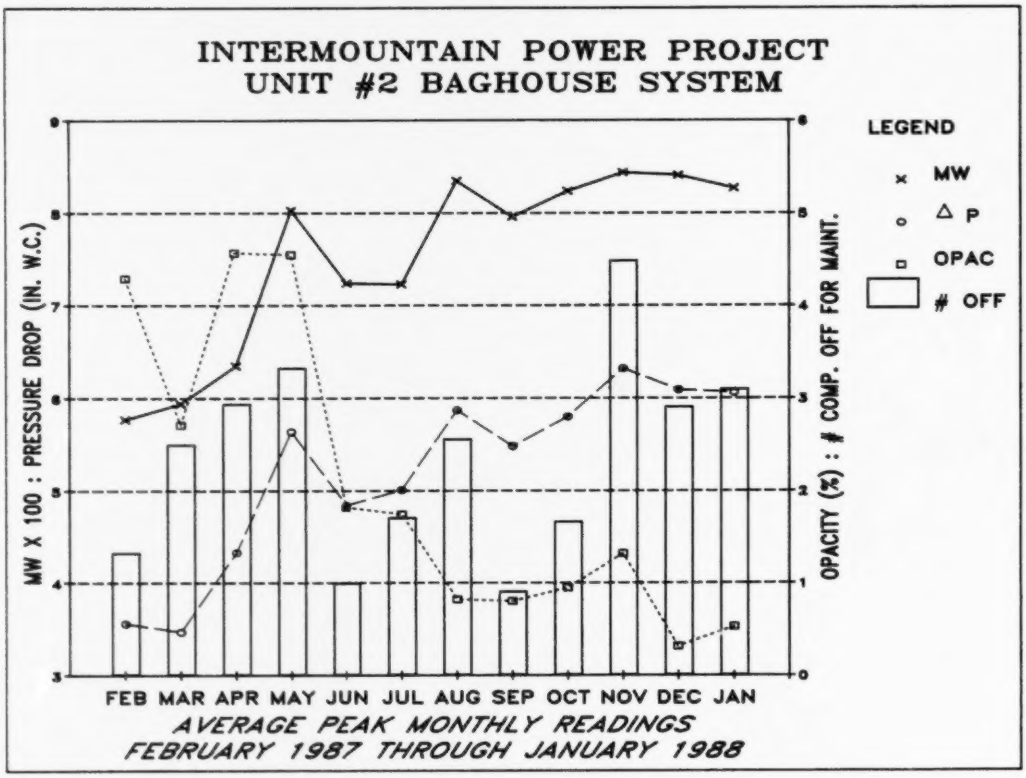


Figure 12. Average peak baghouse system monthly readings.

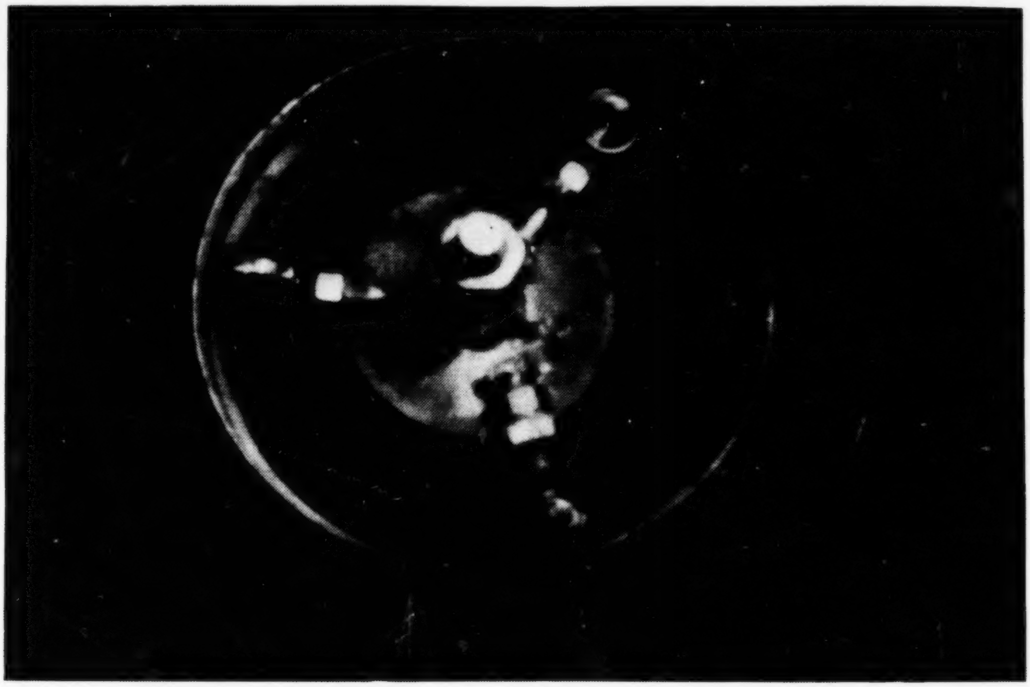
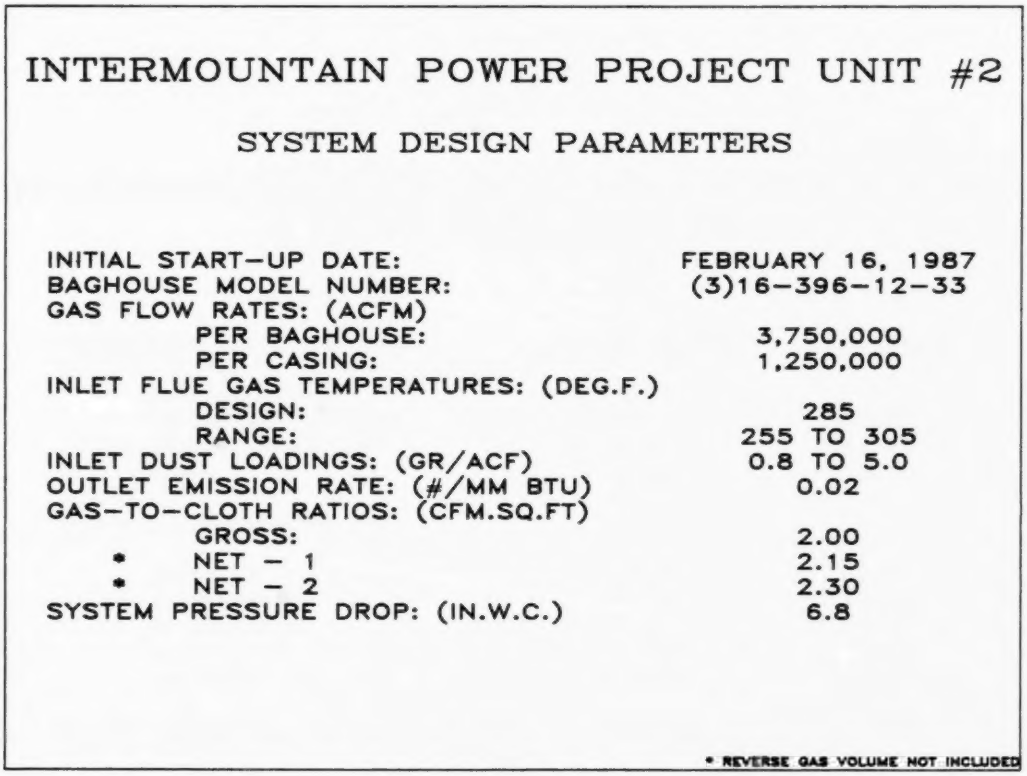


Figure 13. Thimble enlarging device.

TABLE 1. UNIT #2, SYSTEM DESIGN PARAMETERS.



INTERMOUNTAIN POWER PROJECT UNIT #2

SYSTEM DESIGN PARAMETERS

INITIAL START-UP DATE:	FEBRUARY 16, 1987
BAGHOUSE MODEL NUMBER:	(3)16-396-12-33
GAS FLOW RATES: (ACFM)	
PER BAGHOUSE:	3,750,000
PER CASING:	1,250,000
INLET FLUE GAS TEMPERATURES: (DEG.F.)	
DESIGN:	285
RANGE:	255 TO 305
INLET DUST LOADINGS: (GR/ACF)	0.8 TO 5.0
OUTLET EMISSION RATE: (#/MM BTU)	0.02
GAS-TO-CLOTH RATIOS: (CFM.SQ.FT)	
GROSS:	2.00
* NET - 1	2.15
* NET - 2	2.30
SYSTEM PRESSURE DROP: (IN.W.C.)	6.8

* REVERSE GAS VOLUME NOT INCLUDED

TABLE 2. IPP FABRIC TESTING ACCEPTANCE CRITERIA.

FABRIC TESTING ACCEPTANCE CRITERIA

FABRIC PARAMETERS	AVERAGE SPECIFICATIONS
WEIGHT (Oz/Sq.Yd)	12.825 – 14.175
COUNT: WARP	42 – 46
FILL	22 – 26
WEAVE	3 x 1 TWILL
PERMEABILITY (Cu.Ft/Min/Sq.Ft)	35 – 55
YARNS: WARP	37 1/0
FILL	75 1/0
LOI (%)	4.0 MIN.
MULLEN BURST (Lbs/Sq.In)	550 MIN.
TENSILE STRENGTH (Lbs/Sq.In)	
WARP	500 MIN.
FILL	275 MIN.
THICKNESS (In)	0.0137 – 0.0177
WATER DROP ABSORPTION	≤ 1"/30 MIN.
MIT FLEX (CYCLES)	
BEFORE ACID:	
WARP	8000 MIN.
FILL	2000 MIN.
AFTER ACID	
WARP	4000 MIN.
FILL	1000 MIN.

TABLE 3. UNIT #2 TYPICAL FUEL ANALYSIS.

INTERMOUNTAIN POWER PROJECT UNIT #2

TYPICAL FUEL ANALYSIS

PROXIMATE ANALYSIS (%)	AS RECEIVED	DRY BASIS
MOISTURE	5.15	•
VOLATILE MATTER	40.10	42.30
FIXED CARBON	44.80	47.20
ASH	9.89	10.40
ULTIMATE ANALYSIS (%)		
SULFUR	0.48	0.51
CARBON	67.70	71.40
HYDROGEN	4.27	4.51
NITROGEN	1.26	1.33
OXYGEN (By diff.)	11.20	11.90
GROSS HEATING VALUE (Btu/lb)	12010	12660

TABLE 4. UNIT #4 BAGHOUSE PERFORMANCE TEST RESULTS.

BAGHOUSE PERFORMANCE TEST RESULTS

CONTRACT TEST REQUIREMENTS	FORMAL TEST RESULTS JULY, 1987	
	DESIGN	ACTUAL
* MEAN INLET VOLUME: (ACFM)	3,750,000	3,521,523
SYSTEM AVERAGE PRESSURE DROPS: (IN. W.G.)	6.8	4.23
EMISSION RATES:		
(GRAINS/ACF)	0.008	0.0015
(#/MM BTU)	0.02	0.0050
STACK OPACITY: (%)	20	1.5–3.0
UNIT RATING: (MW)	820	836

* AVERAGE BETWEEN PITOT TUBE AND STOICHIOMETRIC VOLUMES

TABLE 5. FABRIC FINISH TEST COMPARISONS.

BAGHOUSE FABRIC FINISH COMPARISONS
UNIT #1 (GE I–625) VS UNIT #2 (I 625–G)

		GE I–625		I–625 G	
FABRIC TESTS EVALUATED		LOW	HIGH	LOW	HIGH
LOSS ON IGNITION:		4.5	5.7	4.6	5.6
PERMEABILITY:		36.2	54.0	41.8	49.6
MULLEN BURST:		565	821	757	901
TENSIL STRENGTH:	WARP	480	579	645	699
	FILL	285	335	362	392
MIT FLEX STRENGTH:	WARP	17,000	41,000	36,000	41,000
	FILL	2,860	5,830	9,639	10,905
ACID FLEX TEST:	WARP	6,130	9,385	12,436	19,918
	FILL	446	1,042	2,920	3,259

MONITORING OF STARTUP AND OPERATION OF INTERMOUNTAIN POWER PROJECT UNIT 1 FABRIC FILTER

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ABSTRACT

The Intermountain Power Agency and the Electric Power Research Institute (EPRI) co-sponsored an evaluation program to develop procedures for optimizing the operation of the Intermountain Power Project (IPP) Unit 1 fabric filter. Southern Research Institute (SRI) conducted the program, documenting and analyzing the operation of the baghouse so that the results could be used at IPP and other utilities to improve baghouse design and operation.

The specific components of the program included augmenting the monitoring system supplied by the baghouse vendor; identifying startup and operational problems; measuring the actual performance against the design or expected performance; relating the performance to boiler operation, fabric material, and fly ash and fuel properties; and developing procedures for startup, operation, and maintenance that would minimize pressure drop and maximize bag life. This paper reports the results of this program, describing the key factors influencing baghouse performance at IPP.

INTRODUCTION

EPRI and the Intermountain Power Agency sponsored an evaluation program to monitor performance and develop procedures for optimizing the operation of the IPP unit 1 fabric filter and to anticipate and eliminate potential problems for the startup and operation of the unit 2 fabric filter. SRI was the contractor for this program. The startup and operation of the unit 1 fabric filter was documented so that it could serve as a model of design and operation for IPP and the utility industry in general. The specific components of the program included the following tasks:

- Provide a high-resolution baghouse monitoring system to augment the instrumentation provided by the baghouse vendor.

- Identify problems and record baghouse performance during precoat, startup, and operation of the baghouse.
- Measure actual performance against design or expected performance.
- Document properties of the bags, fly ash, fuel, and operation.
- Relate the measured performance of the baghouse to the design, boiler operation, fuel and fly ash properties, and bag material.
- Interpret the performance of the baghouse by comparison with the performance of other baghouses in the EPRI/SRI database.

This paper will describe the results of the evaluation of the first 16 months of operation of the IPP fabric filter, from startup on March 28, 1986, through July 31, 1987. The paper gives an overview of problems during startup, problem identification with the EPRI/SRI prototype baghouse diagnostic monitoring system, evaluation of baghouse pressure drop and emissions performance, studies of bag fabric strength and corrosion, and analysis of the properties of coal and dustcake ash.

FABRIC FILTER DESCRIPTION

IPP, located near Delta, Utah, consists of two identical coal-fired units. The coal is delivered to IPP by rail and truck from several mines located in Utah. Units 1 and 2 are each rated for nominal output of 800 MW. The boilers are a Babcock & Wilcox design, each rated at a steam flow of 6,100,000 lb/h at 2640 psi. The turbines and generators were manufactured by General Electric and have a nominal net capacity of 761 MW each.

The air quality control system (AQCS) provides particulate removal downstream of the air heaters by a reverse-gas-cleaned fabric filter supplied by General Electric Environmental Services (GEESI). Flue gas desulfurization is accomplished by wet scrubbers, also from GEESI, installed downstream of the fabric filters.

The unit 1 fabric filter consists of three identical casings in parallel, identified as casings A, B, and C. The arrangement of the casings is shown in figure 1. The air heater outlet duct is divided at the inlet to the fabric filter, with a branch going to each casing. Four induced-draft fans in parallel (from a common baghouse outlet duct) pull the flue gas through the baghouse. A bypass damper in the inlet duct to each casing permits each casing to be bypassed individually.

Each casing is divided into sixteen compartments. There is no provision for purging the entire casing with ambient air; the fabric filter can be purged only through the boiler or by ventilation of individual compartments. Below each compartment is a single hopper for fly ash collection.

There are 396 bags per compartment. The filter bags are 12 in. in diameter and 33 ft long, with an effective filter cloth area of 98 ft^2 per bag and 38,772 ft^2 per compartment. The bags were manufactured by Midwesco using fabric supplied by Burlington. The fabric was required to pass the quality control criteria listed in table 1. The filter bag material has an acid-resistant finish applied before the bags were manufactured. The bags were fabricated with the warp side out (texturized fill yarns on the filtering side). There are eight cadmium-plated rings in each bag to prevent the complete collapse of the bag during reverse-gas cleaning. The bags, which were installed with an applied tension of 75 lb to keep the bags taut, are held on clampless thimbles by compression rings that are sewn into the bag cuffs.

PERFORMANCE GUARANTEES

GEESI's performance guarantee is based on the following operating conditions for the unit 1 fabric filter:

- Maximum continuous flow rate: 3,750,000 acfm
- Emissions rate: 0.02 lb/10^6 Btu; opacity equivalent to 20% at the stack
- Casing pressure drop: 6.8 in. H_2O
- Bag life: 3 years

The air-to-cloth ratio (A/C) at the maximum continuous flow rate is 2 acfm/ft^2 with all 48 compartments in service and 2.3 acfm/ft^2 with one compartment off-line in each casing and one compartment cleaning in each casing.

EPRI/SRI PROTOTYPE BAGHOUSE DIAGNOSTIC MONITORING SYSTEM

SRI installed an EPRI/SRI prototype baghouse diagnostic monitoring system on each casing of the fabric filter to provide enhanced sensitivity in monitoring its operation. The system provides real-time recording of six selected baghouse parameters with three dual-channel strip-chart recorders:

- Gas flow rate through each casing
- Pressure drop across each casing
- Tubesheet pressure drop for each compartment
- Opacity at the outlet of each casing
- Inlet flue gas temperature for each casing
- Temperature for each compartment

Signals from plant instruments (thermocouple transducers, pressure transducers, annubar flow transducers, and Lear Seigler opacity monitors) provide input to the recorder analog interfaces. Recording sensitivity and chart speed can be adjusted to magnify the scale to permit analysis of details that may not be detectable on the less sensitive recorders in the fabric filter control room.

An important feature of the monitoring system is the logic circuitry that detects the status of compartments (on-line, isolated, cleaning, or alarm). When a cleaning cycle is initiated, the logic automatically speeds up the chart from the normal rate of 5 cm/h to 1, 2, or 5 cm/min, and selects the compartment being cleaned to record its tubesheet pressure drop and compartment temperature. Higher speeds are preselected by switches inside the cabinets. Increasing the chart speed on the recorders provides enhanced temporal resolution for easier diagnosis of performance. Manual selection of compartment signals and chart speed can also be made.

RESULTS

The fabric filter was placed in service for the first time on March 28, 1986. Unit 1 has been operated at as much as 900 MW, gross load, and typically generates 840 MW gross under "full-load" operation. The baghouse has operated according to GEESI's specifications with a net A/C of 2.3 acfm/ft^2 and a pressure drop across the baghouse of about 5.7 in. H_2O at full unit load. The residual dustcake areal density has stabilized at 0.26 lb/ft^2.

During the period of this evaluation, the boiler underwent numerous performance tests. Pulverizer operation was changed, excess air was varied, and other conditions were varied over ranges that will probably not be encountered in normal operation. There were numerous outages, especially in the first three months of operation, and one severe boiler tube leak occurred. The baghouse tolerated all of these conditions without an adverse reaction.

The operating logs were used to determine the sensitivity of the baghouse to boiler operation. Specifically, any effects of unit load, the number of pulverizers in service, pulverizer coal flows and pressures, total fuel flow, combustion excess air, oxygen levels in the flue gas, air heater temperatures, ambient temperature, and ID fan currents and flows on baghouse pressure drop and opacity were evaluated. The only correlations between baghouse performance and boiler operating data were to load: an increase in unit load caused an increase in pressure drop due to higher A/C values. Increased boiler load also gives higher ash loadings to the baghouse. The effects of load result in increased frequency of cleaning cycles at high load.

Coals burned at IPP yielded a composite ash that had a low cohesiveness compared to other ashes in the EPRI/SRI database. Specifically, the ash contained a high percentage of calcium (approximately 8 to 9%) and magnesium (approximately 1.9 to 2.5%). These ash properties cause low cohesiveness and result in low dustcake thickness on the bags. Although laboratory tests indicated a relatively low dustcake porosity (approximately 62%), this low porosity coupled with low dustcake thickness resulted in low pressure drop.

Because of the wide range of operating conditions which the fabric filter experienced during the course of this evaluation, it is expected that the long-term performance of the fabric filter, as long as the boiler fuel properties do not change, will remain within the range recorded during the initial 16 months of operation.

PRESSURE DROP

Pressure drops across the three baghouse casings have been controlled by the normal cleaning cycles to less than 6.1 in. H_2O for operation with unit loads up to 900 MW, flue gas flow rates up to 3,960,000 acfm, and air-to-cloth ratios up to 2.9 acfm/ft^2. This pressure drop value is low for the operating air-to-cloth ratio compared to other reverse-gas-cleaned baghouses, as seen in figure 2. The pressure drop has been less than the guaranteed pressure drop limit of 6.8 in. H_2O.

After the first cleaning cycle following startup, the baghouse cleaning cycle was initiated when the pressure drop across a casing reached 5.4 in. H_2O. This has been modified once to increase the limit on casing B to 5.9 in. H_2O, since this casing has the highest gas flow rate. The frequency of cleaning has varied from several hours to about 50 min (continuous cleaning), but has not continued to increase as a function of time-in-service, indicating that dustcake weights have stabilized. The frequency of cleaning is now dependent on boiler load or number of compartments in service.

EMISSIONS

Particulate emission rates from the baghouse were measured during acceptance tests in March 1987 to be between 0.001 and 0.009 lb/10^6 Btu. The guaranteed emission rate is 0.02 lb/10^6 Btu.

Opacity

When compartments are taken off-line for cleaning, the contribution of individual compartments to the outlet opacity can be isolated on recordings from the EPRI/SRI prototype baghouse diagnostic monitoring system. A decrease in the baseline opacity occurring when a compartment is taken out of service, or a precipitous increase in opacity after a compartment that has been isolated is returned to service, are both indicators of bag failures or other leaks in the compartment.

Opacity measured in the outlet ducts of all three casings began to increase dramatically about May 5, 1986. Beginning on May 8, SRI personnel began to systematically check individual compartments to locate any sources of high opacity. The opacity trace from the Lear Seigler transmissometers was monitored while a compartment was isolated. An example trace taken during these tests is presented in figure 3. In this figure, compartments 1C14, 1C15, and 1C16 were each found to be making contributions to overall opacity. Opacity dropped when each of these compartments was taken out of service and increased when each was returned to service. In the case of 1C16, opacity dropped from 9 to 6%. A total of nine bags in compartment 1C16 were found to be off the thimble or had holes in them.

Figure 4 shows the effect of failed bags on baseline opacity. As failed bags were replaced in the compartments, beginning in mid-May, the opacity level decreased to less than 2.5% in all casings.

The baseline opacity at the outlet of the casings was normally less than 2.5%, as shown in figure 5, and is not expected to increase except during bag failure events. This good emissions performance is typical of utility baghouses.

BAGS AND FABRIC

A total of over 415 bags were damaged and replaced during the first year of operation. Of these, 197 could be called startup failures and 164 were due to an operating error in starting up a compartment with the doors open. The startup and subsequent failures have all been related to dimensional tolerances on thimbles or procedural error. There has been no correlation between bag failures and fabric or bag lot number.

Bag fabric samples having up to one year of service were submitted for tests to two laboratories. The fabric breaking strength and permeability of the used fabrics were the only measured properties falling outside the specifications for new fabric. The results of both laboratories were similar: no trend was found between any fabric property and time in service. The bag fabric was not losing strength as a function of time in service, even though the baghouse experienced numerous no-purge outages during the first year of operation and operated with frequent cleaning cycles (due to high unit load) most of the time. Therefore, the bags are expected to meet the guaranteed lifetime of three years and, barring extremely adverse operating conditions, should yield 4 years of service or more.

CORROSION

After 15 months of service, negligible corrosion of any metals in the baghouse had occurred. The baghouse had undergone numerous outages and dew point excursions during this time. The total corrosion that can be attributed to the environment in the compartments was, for the most susceptible metal type, 0.00188 mil, a negligible amount of corrosion.

DUSTCAKE BUILDUP AND ASH PROPERTIES

The bag weights in all casings leveled off to a stable value of approximately 38 lb. Average bag weights by casing are shown in figure 6. Since each bag weighed 13 lb when clean, an average of 25 lb of ash accumulated on each bag. This dustcake weight corresponds to an areal density of 0.26 lb/ft^2. This low value of dustcake areal density (the value generally ranges from 0.5 to 0.75 lb/ft^2 for reverse-gas-cleaned baghouses burning low-sulfur coal) is attributed to the IPP coal and fly ash properties. The low sulfur content of the coal and the high alkalinity (especially the moderately high calcium content [approximately 8 to 9%] and high magnesium content [approximately 1.9 to 2.5%]) of the fly ash at IPP are believed to have played a dominant role in the performance of the baghouse (based on preliminary correlations of performance versus ash properties being developed by SRI under EPRI sponsorship). These ash properties combined with low cohesiveness to yield a thin residual dustcake. This resulted in a low baghouse drag (pressure drop divided by A/C), evident in figure 2.

MECHANICAL PROBLEMS AND MALFUNCTIONS

The number and types of problems that have been observed on the unit 1 baghouse are not excessive for new equipment of this size and complexity. Malfunctions have been reported in outlet, reverse-gas relief, and inlet dampers. Some of these damper problems prohibited compartments or casings from being cleaned, which could have led to high pressure drop, high bag weights, and shortened bag life if unattended. The purge air fans failed. The ash disposal system had serious erosion problems in the pressure-equalizing line in the compartment hoppers directly beneath the tubesheet, which led to bags being destroyed by ash from the pressure-equalizing line spraying through thimbles onto the inside of bags. The Gould-Modicon programmable controller has erred in the timing of cleaning cycles. The most severe mechanical problem was undersized bag thimbles, which led to many bag failures when bags popped off these thimbles. However, all of these problems were either corrected as part of startup or later as part of maintenance.

EARLY DETECTION OF PROBLEMS

The most significant feature of the research project, especially in light of the good operating performance of the unit 1 baghouse, was the use of the EPRI/SRI prototype baghouse diagnostic monitoring system to identify operating problems rapidly. The system's ability to change recording speed and scale provides the resolution necessary to identify and locate problems with fans and dampers, errors in the controller cleaning cycle sequence, and bag failures. These problems can be diagnosed quickly before serious equipment damage, high pressure drops, high emissions, or major maintenance and bag replacement costs occur.

CONCLUSIONS

Attention to baghouse operation, using the EPRI/SRI prototype baghouse diagnostic monitoring system, will permit early detection of problems in dampers, bags, and cleaning effectiveness before they become severe. The baghouse diagnostic monitoring system was used to:

- Locate bag failures before they propagated.
- Detect damper chatter or faulty actuators before damper or fan wear was severe.
- Identify programming errors that prevented effective cleaning of all compartments.

The following conclusions have been drawn about the fabric filter performance from the results of the monitoring program:

- The baghouse performed within the guaranteed limits of pressure drop and particulate emission rate during this initial 16-month period of operation.
- Boiler upsets, outages, ambient conditions, and variations in the operation of plant equipment did not adversely affect baghouse performance. The only correlations between baghouse performance and boiler operating data were to unit load: an increase in unit load caused an increase in pressure drop.
- The baghouse experienced a low dustcake weight, which resulted in relatively low pressure drop for the operating A/C value. This low dustcake weight is believed to be due to the relatively high percentages of calcium (approximately 8 to 9%) and magnesium (approximately 1.9 to 2.5%) in the ash.
- Based on results of fabric strength tests and the operating history of the baghouse during this evaluation, the bags are expected to meet the guaranteed lifetime of three years and, barring extremely adverse operating conditions, should yield four years or more of service.
- There was no significant corrosion in the baghouse during the 16-month test program, in spite of numerous outages without compartment purging.

The performance guarantee by the baghouse suppliers is a pressure drop of 6.8 in. H_2O. Present operation is, conservatively, 0.7 in. H_2O below this value, which results in an annual savings in I.D. fan power consumption of $105,000 (based on the conservatively calculated cost of $150,000 per year for each inch of pressure drop). Operation of the baghouse at a pressure drop significantly lower than the supplier's guarantee in our evaluation is the result of the following: a) attention to the design and installation of the baghouse system, b) careful selection and quality control of the bag material/finish and bag manufacture, c) fuel properties which create an ash that easily sheds from the bags, and d) the use of the EPRI/SRI prototype baghouse diagnostic monitoring system. The monitoring system enables the IPP plant personnel to detect operational problems (such as faulty dampers and improper operation of the Gould-Modicon programmable controller), which if left uncorrected, would result in baghouse pressure drop increase. Projecting a bag life of four years instead of the guaranteed 3-year bag life would result in a $250,000 capital expenditure saving every fourth year. Careful attention to baghouse operation and performance can maintain these cost savings. The monitoring system is ideally suited for these diagnostic functions, providing information that is difficult or impossible to obtain from other baghouse control room instruments.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

ACKNOWLEDGMENTS

This work was funded under the EPRI Research Project 1129-14. The data were taken by numerous employees of SRI and Intermountain Power Service Corporation. The emissions performance test was conducted by Mostardi-Platt Associates.

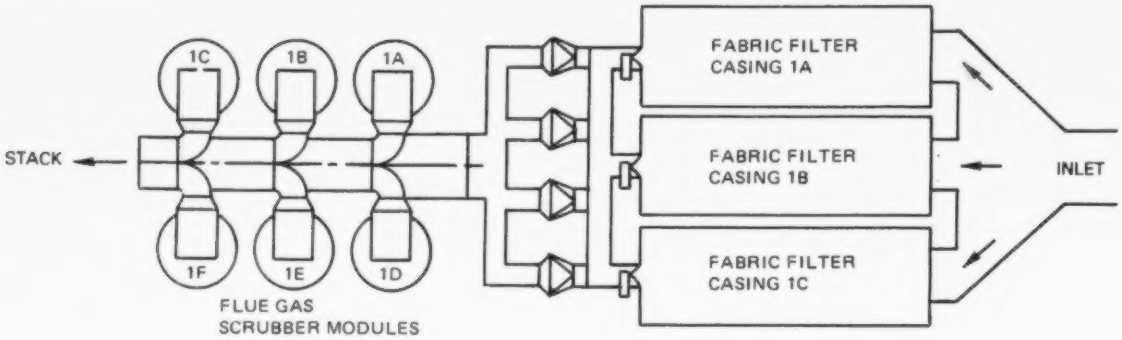

Figure 1. Arrangement of fabric filter casings at IPP unit 1.

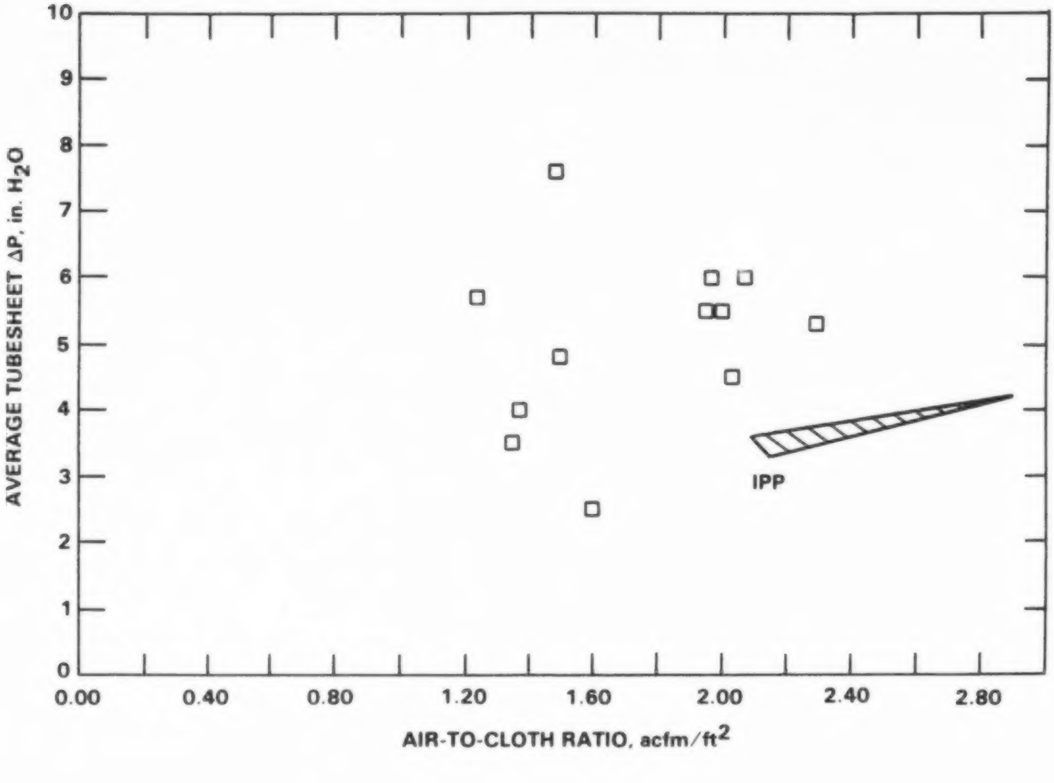

Figure 2. Average tubesheet ΔP recorded in the three casings at IPP January 9-11, 1987, coincident with velocity traverse measurements. Data from 12 other baghouses in the EPRI database are shown for comparison.

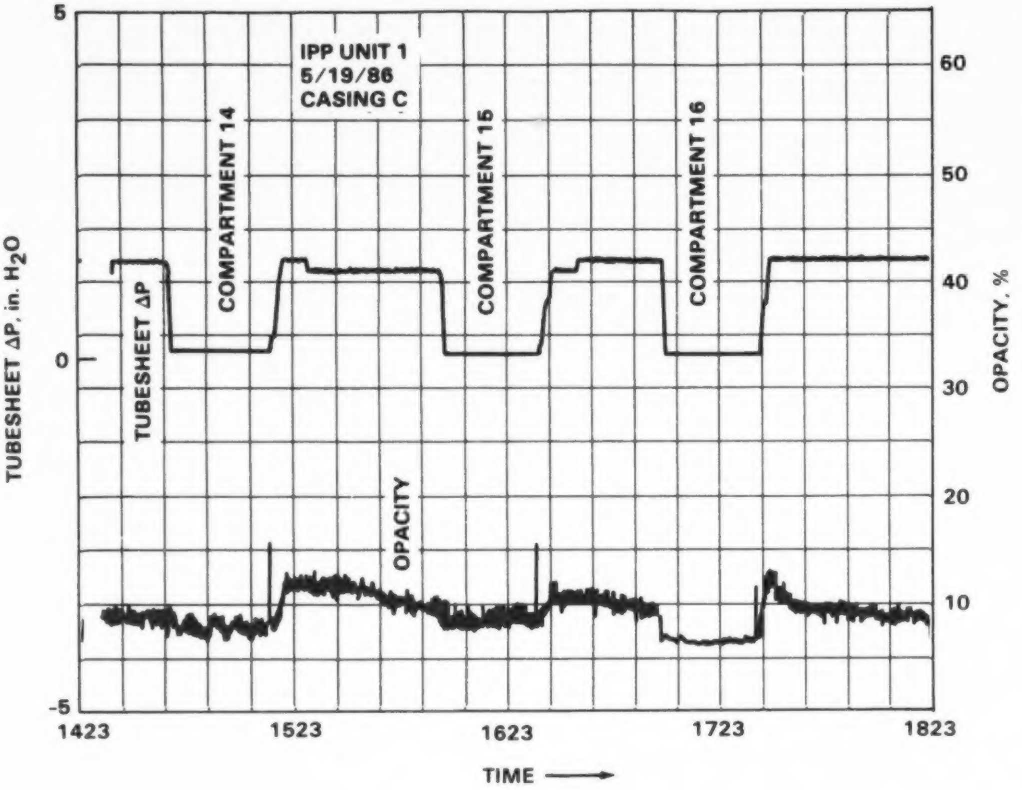

Figure 3. Change in opacity as three successive compartments were taken off-line.

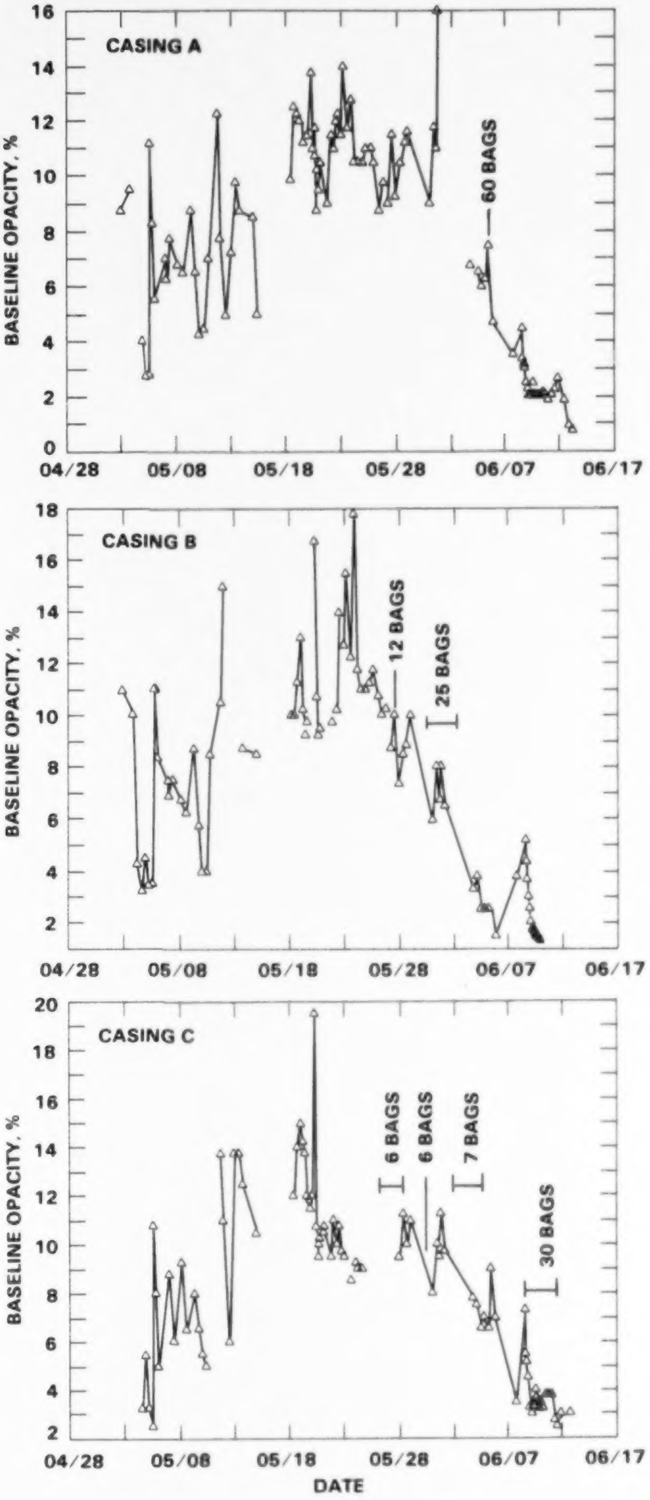

Figure 4. Baseline opacity for each casing. The numbers of bags indicate when groups of failed bags were replaced.

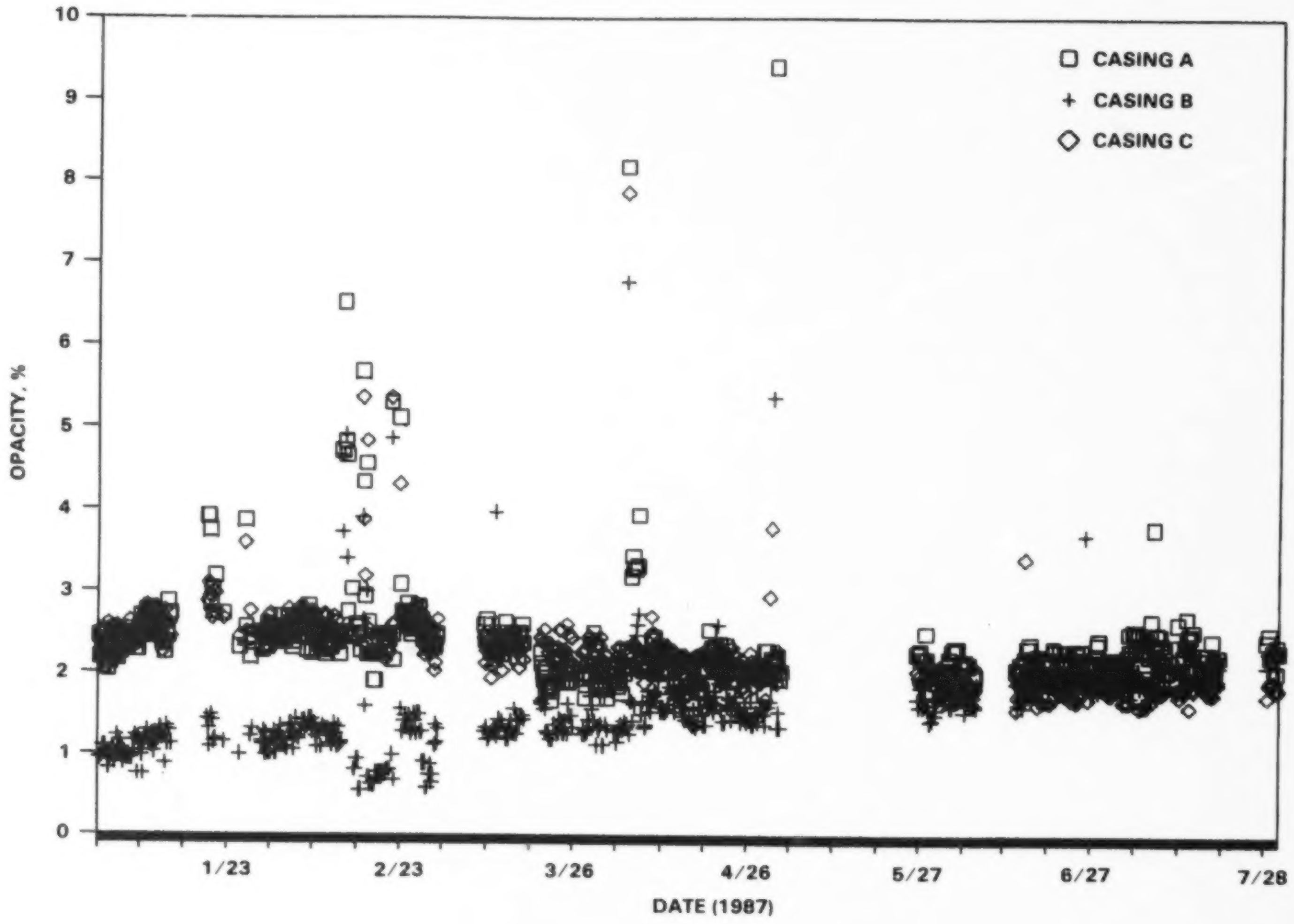

Figure 5. Opacity values recorded for each baghouse casing during the period from January 1 through July 31, 1987. Opacity excursions above the baseline levels of 1 to 3% are due to bag failures.

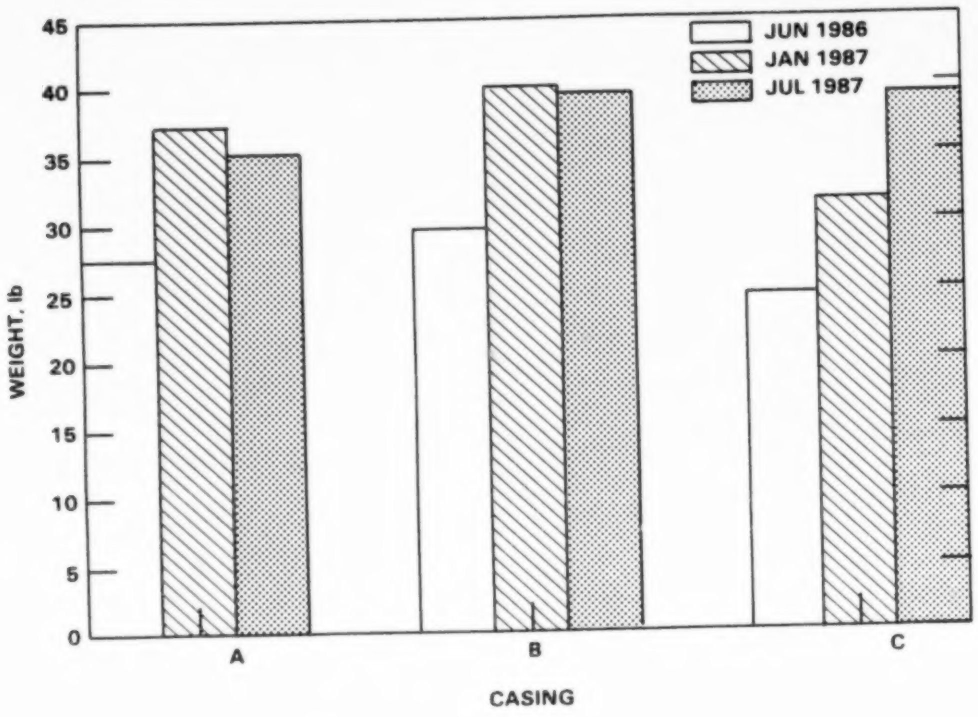

Figure 6. Average bag weights by casing for June 1986, January 1987, and July 1987.

TABLE 1. FABRIC SPECIFICATIONS FOR IPP

Weight, oz/yd^2	13.5 + 5%
Construction	3 x 1 twill
Thread Count	Warp: 44 + 2
	Fill: 24 + 2
Texturization, %	75
Thickness, in.	0.00157 + 0.002
Loss on Ignition, %	4.0 - 5.5
Permeability, cfm/ft^2 at 0.5 in. H_2O	45 + 10
Mullen Burst, lb/in^2 minimum	550
Breaking Strength, lb/in^2 minimum	Warp: 500
	Fill: 275
MIT Flex, cycles minimum	Warp: 8,000
	Fill: 2,000

FABRIC FILTER EMISSIONS,
STUDY OF THEIR VARIABILITY

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ABSTRACT

The fabric filter has taken a significant portion of the electric utility emission control market. One of the reasons has been the belief that emissions from fabric filters are almost always low and are more predictable than emissions from electrostatic precipitators.

A data base of over 50 operating fabric filters indicates wide variation of emissions. This paper presents part of the data base and examines the influence of design and operating variables on particulate emissions. Parameters possibly affecting emissions include air to cloth ratio, type of fabric filter (reverse air or pulse clean), bag length, bag attachment method, fabric weight, fabric type, inlet particulate concentration, year of startup, gas temperature, ash characteristics, and plant size. Unfortunately, due to the many design and operating variables at each plant, it was not possible to create accurate correlations. However, several interesting conclusions are reached.

A discussion is presented of what can be incorporated into the design and operation to help assure very low emissions and what variables affecting performance need further investigation.

INTRODUCTION

In 1984 Mr. A. Khosla and I presented a paper (1) comparing the performance of fabric filters and electrostatic precipitators on total particulate, fine particulate and trace element emissions. The paper included a data base of 20 fabric filters and 18 well-designed precipitators. The paper concluded that the average total particulate emissions of the operating fabric filters was about the same as the well-designed precipitators, but that the fabric filters were superior for fine particulate and trace element emission control.

Since 1984, the fabric filter data base has grown from 20 to nearly 60 units. The fabric filter emissions of the operating units continues to vary over a wide range. The purpose of this paper is to examine the data base to determine if the data indicate reasons for this large range of performance.

Recently we have been asked by several state agencies to design for emission levels well under the new source performance standards. It is no longer a system of "can we comply with the EPA new source performance standards". Sometimes we hear "well, Smith Unit 2 has emissions of 0.001 grains per cubic foot so we expect you to do as well". If a plant starts up and achieves emission levels far below what was anticipated, due to some quirk of luck or combination of circumstances not understood by science, it cannot be expected that the next one can be designed and guaranteed for as low of an emission rate. The range of data and analysis presented in this paper will show that design of a fabric filter for ultra low emissions is not a well understood science, its not even a black art.

THE FABRIC FILTER DATA BASE

Table 1 shows a summary of the emission data from the design and performance data base. The data base consists of information which has been published about each fabric filter. We are grateful for the efforts of a great number of individuals who have shared their knowledge by preparing technical papers to help the industry. Each emission data point on Table 1 is referenced to its source. The table shows a small portion of the total data base which includes design and operating data for each installation.

Not all literature references show emissions in terms of both pounds per million Btu and grains per actual cubic foot. For these plants, an estimate of a conversion to the missing emission unit has been made. This conversion is affected by operating excess air and plant altitude. With these data missing for some plants, the conversion may be slightly high for some while low for others. However, the data averages should be very representative.

The data from each fabric filter should not be interpreted as what occurs daily at each plant. It is expected that emissions vary from month to month. Some plants showing high emissions have been improved. For example, the literature shows rather high emissions for the W. M. Samis Station. These emissions (within the plant's regulatory limit) probably have subsequently been reduced because the reported emission data were from a period when the fabric filters were experiencing severe operating difficulties. On the other hand, some of the other units may have been "tuned up" for the emission tests and emissions may have increased since. However, it is hoped that by use of all the data, the overall influence of design and operating conditions could be seen. The following sections examine these design and operating variables.

TOTAL EMISSION VARIATION

The average emission from the fabric filters is 0.021 pounds of particulate per million Btu. This corresponds to 0.0068 grains per actual cubic foot. Table 1 indicates a considerable range of emissions. Figures 1 and 2 illustrate this wide range of emissions. The data are arranged in ascending order. Separate figures are shown for emissions in terms of pounds/million Btu and grains/acf.

The data indicate that 80 percent of the operating fabric filters are performing better than the electric utility new source performance standard of 0.03 pound per million Btu. In that the compliance is not 100 percent, may be a surprise. However, some of the units do not need to comply with this emission standard. Thus, it is not meant to imply that certain plants have emission rates that are too high. However, the question remains regarding why there is such a large variation of emissions. The fabric filter, unlike an electrostatic precipitator, cannot be designed for a given emission level or percent particulate removal. The units experiencing unusually high or low emissions were not designed that way. There must be other reasons for the emission variation.

If one were to be faced with a regulatory agency requesting design for very low emissions, the data could be used to show the probability of success. For example if one had to design for an emission rate of 0.02 pounds per million Btu, Figure 1 would indicate that only about 40 out of 57 units are achieving it. The chances of success would be only about 70 percent, not very comforting. Engineers like to design with higher confidence levels. Thus, one may eventually face decisions regarding how to design fabric filters for lower emissions.

HISTORIC TREND OF EMISSIONS

Figure 3 shows the trend of emissions from fabric filters starting up during the indicated years. It is not known whether emissions have increased or decreazsed for each unit since the emission data were reported. However, the group of "typical" data in the lower portion of the figure indicates that emissions have been creeping up with time as new units come on line. In 1984, when the previous survey (1) was completed, the average emission from the operating fabric filters was 0.0040 grains/acf. This data base consisted of the first 20 units shown on Table 1. Now, in 1988, the average emission from all listed fabric filters has increased to 0.0068 grains/acf. Again, this is not the effect of ageing. These are emission data from new units.

DESIGN TRENDS WITH TIME

Figure 4 shows that fabric filters have been applied to increasing sizes of plants over the years. The number of bags per compartment has been increasing (Figure 5). However, the length of bags generally has not been increasing (Figure 6). Also, the ratio of bag length to diameter has not changed much with time (Figure 7).

EFFECT OF DESIGN ON EMISSIONS

NUMBER OF COMPARTMENTS AND BAGS PER COMPARTMENT

Figure 8 indicates that the number of compartments does not affect the emissions for the group of performance data. Figure 9 may indicate a slight trend of increasing emissions when the number of bags per compartment increases. Increasing the number of bags per compartment may affect willingness to do maintenance and it may also affect the ease of achieving uniform gas distribution.

FABRIC FILTER SIZE

Emissions have increased slightly as the overall capacity of the fabric filter has increased (Figure 10). Some have speculated that, as fabric filters were applied to larger units, the maintenance tasks would overwhelm the plant staff. However, it appears that this has not been a strong trend.

AIR TO CLOTH RATIO

Figure 11 indicates that there is not a strong correlation of air-to-cloth-ratio to emissions. This even appears to be the case when one focuses only on the units designed for relatively low ratios and operating with low emissions (Figure 12).

Some of the literature references do not report the gas flow rates and number of out of service compartments during testing. This causes difficulty in estimating operating air to cloth ratios. Design values were used when actual operating values were not reported. This lack of accurate information may be a partial cause for the scatter on Figure 11. Another potential cause of the scatter is the unknown frequency of cleaning cycles. It has been reported that adjustments to cleaning frequencies have greatly affected emissions at some plants, reduced cleaning frequency being beneficial.

It would be reasonable to conclude that a lower air to cloth ratio, lower velocity through the fabric, would reduce pressure drop and leakage through holes and would also allow bridging of ash over holes. Laboratory tests at the University of Tennessee (2) indicated that emissions more than tripled when the air to cloth ratio doubled. Other laboratory tests also showed an increase of emissions (3). Tests by DuPont (4) on simulated pulse-jet operation also confirmed the benefit of lower air to cloth ratio on emissions.

How does all this laboratory work compare to field experience? Data has been published for two plants where the ratio was varied by testing at different gas flow rates. These data should be beneficial because they do not include all the variables influencing Figures 11 and 12. Figure 13 shows that emissions increased at the Kramer Station as gas flow and air to cloth ratio were decreased (5). Data from the Parish Unit 8 (Figure 14) also shows this trend (6). Tests were conducted at Parish on four separate parallel fabric filter trains for Unit 8. One train showed wide scatter. The other three trains all showed the same trend, as shown on Figure 14.

At reduced gas flow and boiler load, the quantity of ash entering the fabric filter decreases. The velocity through leakage holes decreases, but the emissions increase. It may be a quirk, or, it could be because the cleaning cycle is not optimized at reduced loads. It is not known whether cleaning frequency was determined automatically by pressure drop increase or set for a given time interval. Perhaps cleaning was removing too much of the ash from the fabric causing more leak-through. Less frequent cleaning at reduced load may reduce emissions.

It is believed that decreased air to cloth ratio would be beneficial in reducing emissions, based on the above referenced laboratory data; but, more field work needs to be done on full size units to demonstrate this.

BAG LENGTH TO DIAMETER RATIO

As the ratio of bag length to diameter increases, more flue gas must pass through the opening at the inlet. This should increase wear, erosion, and lead to premature bag failure and increased emissions. This may be true over wide ranges of length to diameter ratio, and over longer periods of operation, however, the narrow range of data in the data base leads to no conclusions (Figure 15).

GAS TEMPERATURE

Flue gas temperature may affect emissions. Some have demonstrated that certain ashes become more cohesive at reduced temperatures, or under the influence of more humidity with water spraying. The ash would then tend to bridge over and not leak through pin hole openings. Unfortunately, this theory is not supported by the bulk of the data base, as showm on Figure 16. It is not even supported by the data from a single brand of fabric filter, some of whose units have spray dryers upstream, reducing gas temperature (Figure 17). However, recently completed pilot tests, partially sponsored by Bechtel and Ahlstrom, indicated a benefit in water spraying ahead of the fabric filter to reduce temperature (Figure 18).

TYPE OF FABRIC FILTER

Data base emissions from the three types of fabric filters are as follows.

Average Fabric Filter Emissions

		Emissions	
Type of Cleaning	Number of Units	Grains/acf	Lbs/mmBtu
Reverse Air	41	0.0060	0.019
Shake-Deflate	6	0.0058	0.023
Pulse-Jet	9	0.0100	0.031

The data indicate that the pulse-jet type cleaning fabric filters have considerably higher emissions than the other types. The shake-deflate type has a slightly higher emission on a pounds per million Btu basis than the reverse air type. The difference in the above ranking, based on outlet concentration and on pounds per million Btu, is due to the shake-deflate design having three units at high altitude where a given particulate concentration translates into a higher number of pounds per million Btu.

An EPRI study concluded that "In short term testing, emissions from shake-deflate cleaned bags, while well within applicable regulations, were approximately three times higher than from reverse gas cleaned bags" (7). These were carefully controlled tests, however, the magnitude of their results are not supported by the above field data. It is noted that a data base of only six shake-deflate units is not nearly as statistically significant as the much larger reverse air clean data base.

It is interesting to note that one of the claimed advantages of the shake-deflate type cleaning is its superiority in cleaning and keeping the pressure drop down. One of the answers to pressure buildup for reverse clean fabric filters has been the installation of sonic horns. The previously mentioned EPRI study also concluded that "Particulate emissions from reverse gas/sonic cleaned bags appear to be approximately equal to emissions from bags cleaned by reverse gas alone". Thus, if horns had been installed on any of the reverse air clean systems, the emissions may not have been any higher and the above ranking would remain.

BRAND OF FABRIC FILTER

It was thought that some of the brand-dependent aspects of fabric filter design might affect the emissions. These may include method of bag attachment to the tubesheet, compartment inlet design, etc. The following is a summary of emission data for each of the brands having at least three operating fabric filters.

Average Emissions for Various Brands

Brand	Number of Units	Emissions Grains/acf	Emissions Lbs/mmBtu
A	13	0.0029	0.013
B	8	0.0044	0.011
C	3	0.0047	0.014
D	9	0.0057	0.017
E	6	0.0058	0.023
F	4	0.0200	0.068

The wide range of emissions points to the possible conclusion that emissions depend on brand design features. This, however, may be a false conclusion. The study of brand features and their influence on emissions is beyond the scope of this paper.

FABRIC TYPE

A detailed study of the influence of the type of fabric and its manufacturing technique is also beyond the scope of this paper. There are dozens of differences in bag specifications which will influence emissions. For example, some have found that increasing fabric texturization from 25 percent to 100 percent reduces emissions by over 50 percent (8). However, it is not known whether this conclusion, for one set of tests at one plant, is universally applicable. Fabric differences probably are the reason for some of the emission variation in our data base.

It is not possible to reach conclusions based on the data bank information because the overwhelming majority of fabric filters of the reverse air clean type use fiberglass bags. Thus, comparison to other fabric materials is not possible. Furthermore, their weave, texturization, manufacturing technique, etc. generally are not stated in the literature.

FABRIC WEIGHT

As with fabric type, there is insufficient information in the data bank to reach conclusions. Most of the reverse air clean fabric filters use 9.5 to 10.5 ounce fabric. Only a few use 13 ounce fabric and none reportedly uses 16 ounce fabric. For several of the units, the weight of fabric has not been reported. Thus, data correlations showing the effect of fabric weight on emissions are not possible.

INLET ASH CONCENTRATION

The fabric filter has a reputation for producing emission particulate concentrations which are not dependent on inlet concentrations. However, the effect of leakages through holes in the fabric and leakages around the fabric should produce increasing emissions as fuel ash content increases. Various laboratory tests have confirmed this. Also, more frequent cleaning, due to higher fuel ash content, leads to higher emissions. One could, with some imagination, come to this conclusion based on the data bank information reflected in Figure 19. Unfortunately, because of the scatter of other variables, a strong correlation is not noted.

To design for ultra-low emissions, the effect of fuel ash content and the amount of leakage and bypass gas are important factors needing further investigation. It appears that a lower ash content would reduce emissions by reducing leakage and also by allowing less frequent fabric cleaning.

FUEL AND ASH PROPERTIES

Several papers have recently been presented showing that fuel properties affect the emission level. Laboratory research has indicated that increasing amounts of sodium, phosphorous and magnesium in the ash may reduce emissions (9). A study of the ash constituents and their influence on emissions has not been attempted for this paper. However, one variable which was studied is the coal sulfur content. The available information does not show a correlation of fuel sulfur to emissions.

Ash shape will also affect emissions. Fuel and combustion properties at the Monticello Station have produced slippery round spheres of ash which bleed through the fabric more readily than most ash. On the other hand, ash produced in fluidized bed combustors is less likely to penetrate fabric because of its jagged geometry, due to lower furnace temperature. The area of fuel properties and combustion characteristics needs further investigation.

GAS CONDITIONING

Gas conditioning has been used extensively for aiding the performance of electrostatic precipitators. It thus is of interest that gas conditioning has been found to benefit fabric filter emissions. Tests at the Monticello Station (8) indicate a more than 50 percent reduction of emissions with use of ammonia gas conditioning. This was observed for several different fabrics. It is possible that the ammonia makes the ash more cohesive and therefore less likely to bleed through the fabric. The Monticello ash has been reported to have a unique bleed through characteristic. Thus, the ammonia conditioning test results may not be applicable to all plants.

Laboratory tests in North Dakota (9) indicate dramatic reductions of emissions for certain conditions and ashes with use of ammonia and sulfur trioxide gas conditioning. It is possible that some of the scatter of emissions in our data bank is due to naturally occurring gas conditioning, i.e., variations in flue gas moisture, variation in conversion of sulfur dioxide to sulfur trioxide among boilers, etc. Flue gas conditioning is an interesting area needing further testing to see if it is beneficial to conditions other than those at the Monticello Station. It is noted that new units in California are being required to install ammonia injection systems for NOX control. It is possible that unreacted ammonia may have a positive side benefit, increasing particulate removal in the fabric filter.

COMBINED EFFECTS

The previously discussed graphs do not paint a very positive correlation of design and operating parameter influences on emissions. Part of this may be due to the vast number of inter-related variables. Analysis of variable combinations may eventually show better correlations. For example, if it were postulated that both increasing air to cloth ratio and increasing bag length (more erosion potential) would increase emissions, a correlation of the mathematical product of these variables to emissions would be useful. Figure 20 shows this correlation. It is possibly significant that the only units having emissions higher than 0.02 pounds per million Btus, also have bag length-times-air-to-cloth-ratio indices of over 55. This is but a first attempt at inter-related parameter correlation.

CONCLUSIONS

The data from operating fabric filters indicate that a wide range of emissions can be expected. It is difficult to use the data to show that specific design differences influence emissions. This is because of the existence of too many design and operating variables. One could conclude that fabric filter design is rather forgiving, wide ranges in design do not predictably affect emissions. This, however, is believed to be a false conclusion. If one could spend enough time and money pilot testing variables, conclusions could be reached regarding their influence on emissions.

It is not thought that the industry has the financial capability for the level of testing which would be required to reduce fabric filter design from that of a black art to that of a science. Therefore, it is of interest to continue to develop correlations from expanding data beses and laboratory tests.

Differences in design and operation influencing emissions probably include most of the following.

1. Fabric leaks
2. Bypass damper leakage
3. Cleaning frequencies and techniques
4. Fuel and combustion characteristics
5. Maintenance quality
6. Fabric weight and design
7. Height to diameter ratio
8. Air to cloth ratio
9. Gas temperature, humidity and SO3 content
10. Fabric filter brand-specific features

In cases where emissions must be designed for levels considerably below the utility boiler new source performance standards, 0.03 pound per million Btu, it is advisable to take special precautions in the fabric filter design. These may include the following.

1. Elimination of bypass duct or use of double bypass dampers with purge air.
2. Use of a lower air to cloth ratio combined with less frequent cleaning.
3. Selection of specific fabric materials, types and construction.
4. Possible use of gas conditioning, if proven in more tests at other plants..
5. Use of reverse air rather than pulse-jet cleaning.
6. Possible use of a reduced temperature by revising the air preheater design or installation of a spray dryer upstream of the fabric filter. This would be especially helpful in jurisdictions which include condensable material as particulate. Some of the condensed material would be removed by the fabric filter at reduced temperature.

Emissions may also be reduced by putting more effort into maintenance. One could speculate that the plants having higher emissions may have less maintenance, but this would be an unfair accusation against some of these plant staffs. There simply is not much data regarding maintenance level correlation to emissions.

The above list of design features for low emissions are tentative conclusions. Each represents an area where more research is needed. Much of the research which has been done is aimed at areas other than how to achieve lower than new source performance standard emissions. However, there are many cases where it is necessary to go to lower emission levels. With the average of all emission data showing levels two thirds as high as the new source performance standards, there is little room for error. If, for example, we had to design for one third lower emissions, 0.02 pound per million Btu, there would only be a 70 percent chance of achieving it, based on experience at 57 plants. This is not a comfortable design margin.

Some of the problems of operating data interpretation would be lessened if there was a test and reporting standard listing what information would be useful. So often we see test results and wonder, if only they had taken fuel samples or if only they had noted how many compartments were out of service or what the cleaning cycle was. Documentation of operating conditions when reporting test results would be helpful in developing performance correlations.

ACKNOWLEDGEMENTS

The author wishes to express appreciation to the 36 authors, and their numerous coauthors, listed in the references. It is the hard work of these people who have reported design and operating data which made this data analysis possible. Special thanks go to Mr. Walter Piulle and EPRI who have provided so many good data bases over the years, including the referenced (32, 33 & 34) fabric filter data bases.

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The work described in this paper was not funded by the U.S. Environmental Protection Agency. The contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

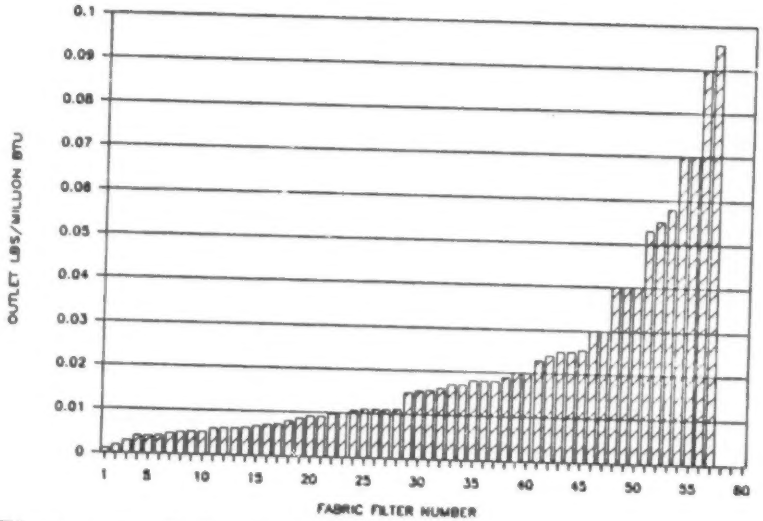

Figure 1. Emissions, Lbs./mmBtu, 57 plants.

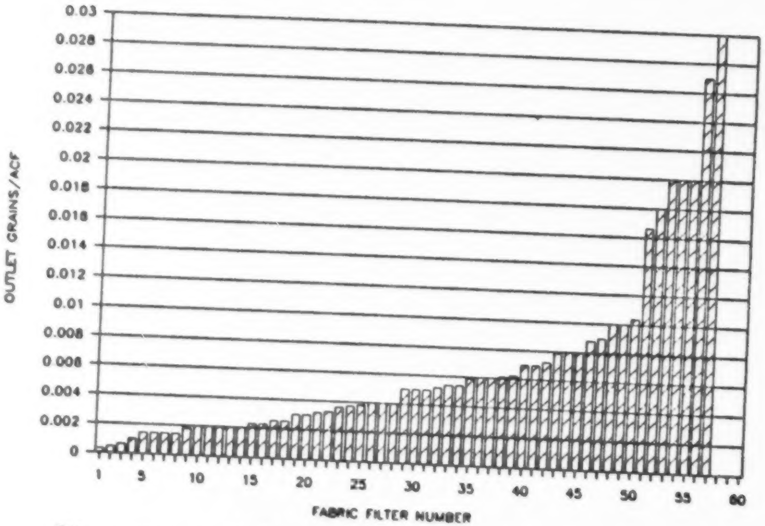

Figure 2. Emissions, grains/acf, 57 plants.

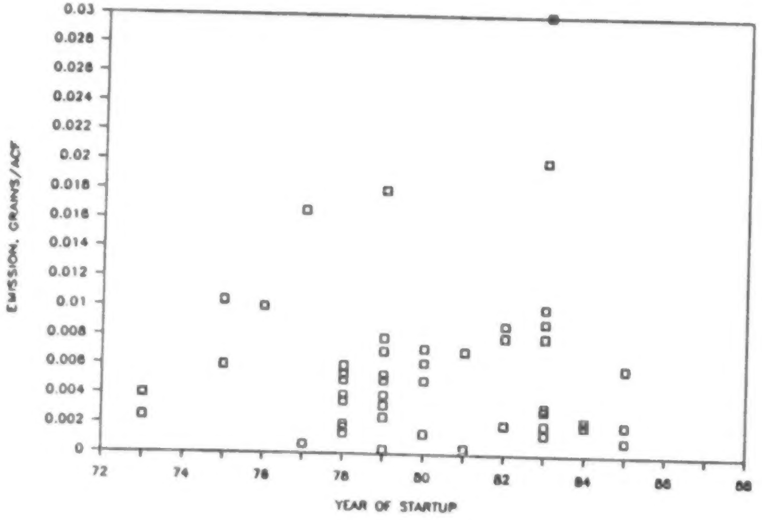

Figure 3. Emissions versus year of start-up.

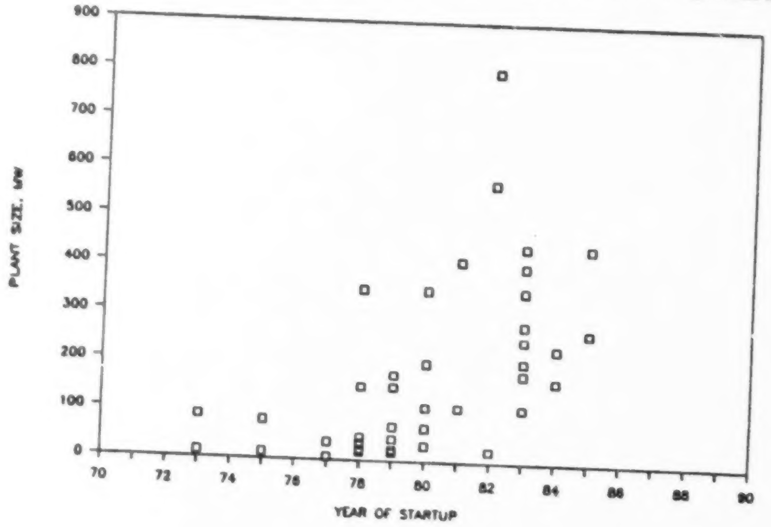

Figure 4. Fabric filter plant sizes.

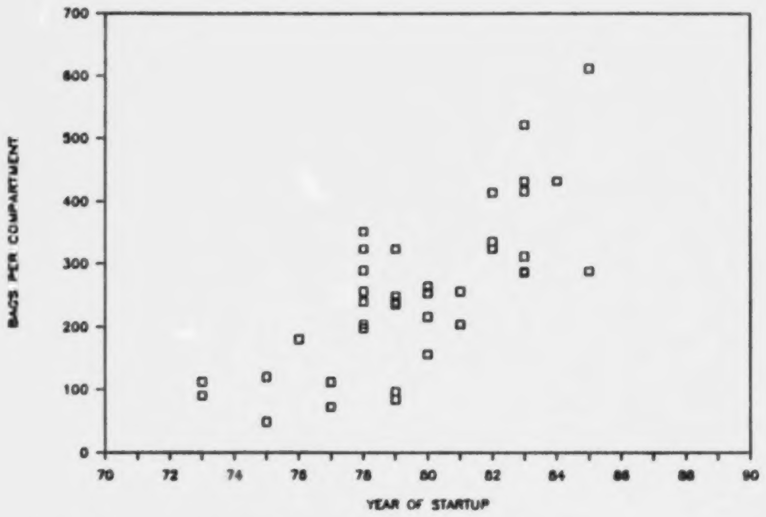

Figure 5. Compartment size growth.

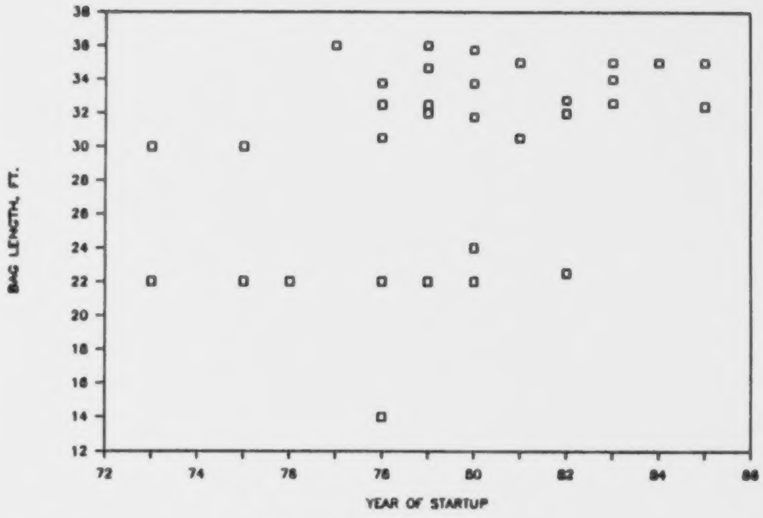

Figure 6. Bag length history.

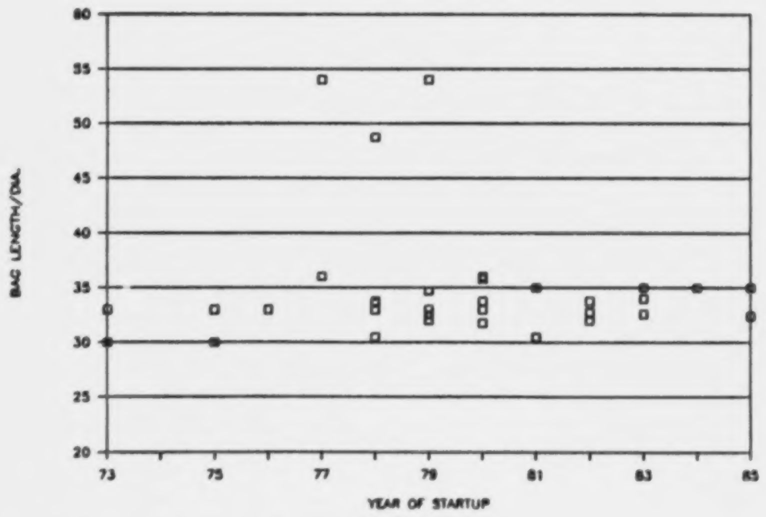

Figure 7. Bag length/diameter.

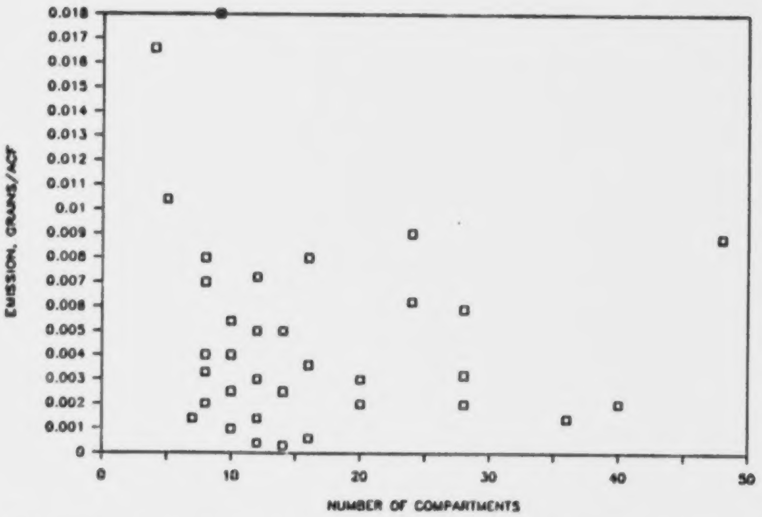

Figure 8. Emissions vs. number compartments.

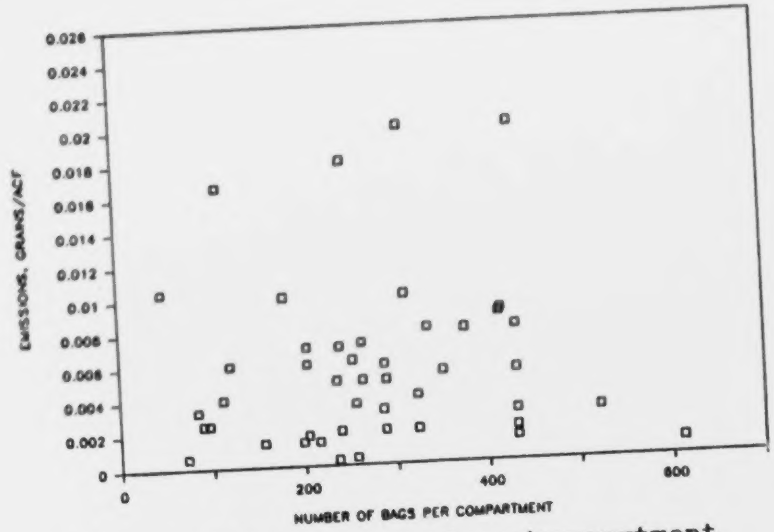

Figure 9. Emissions vs. bags/compartment.

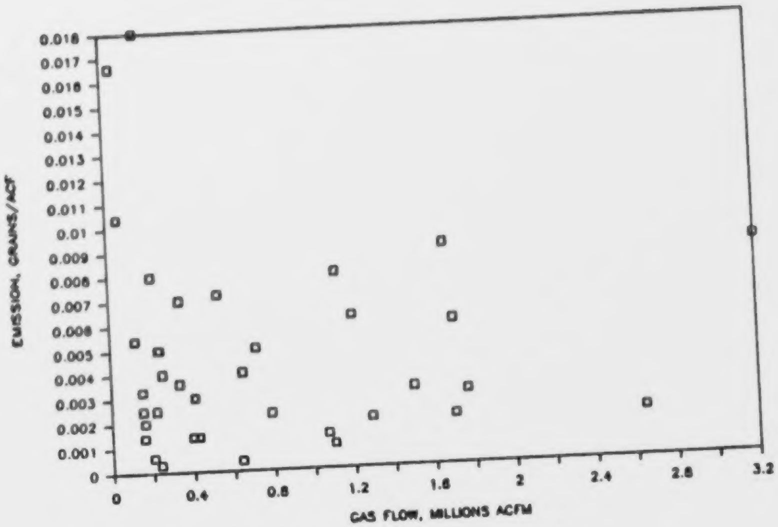

Figure 10. Emissions vs. gas flow.

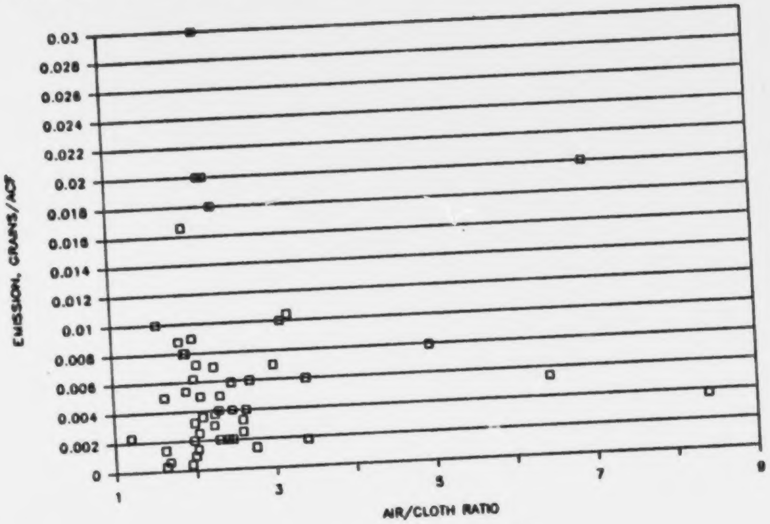

Figure 11. Emissions vs. A/C,
all reverse air units.

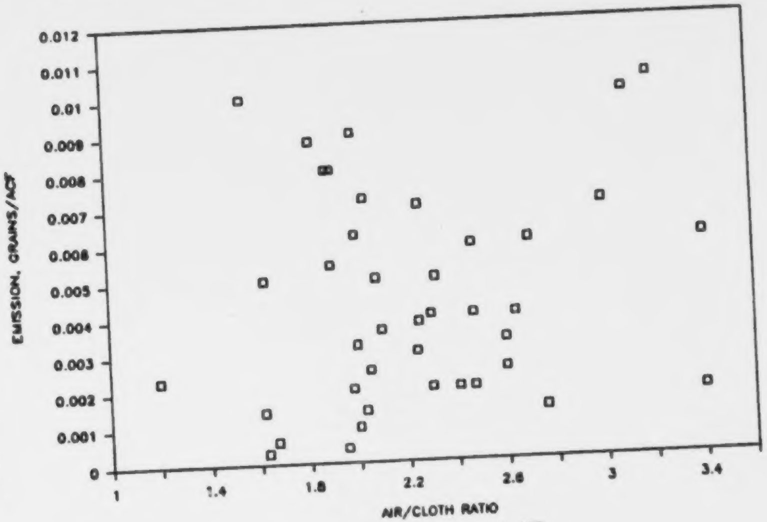

Figure 12. Emissions vs. A/C,
low emission units.

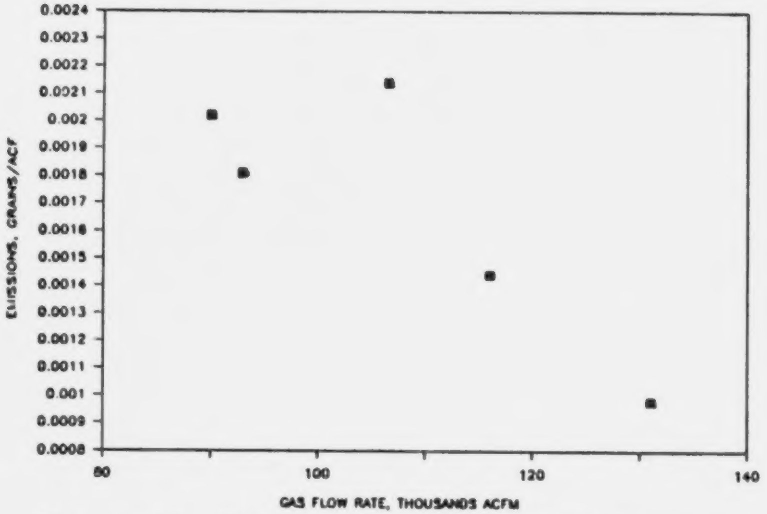

Figure 13. Emissions vs. gas flow, Kramer.

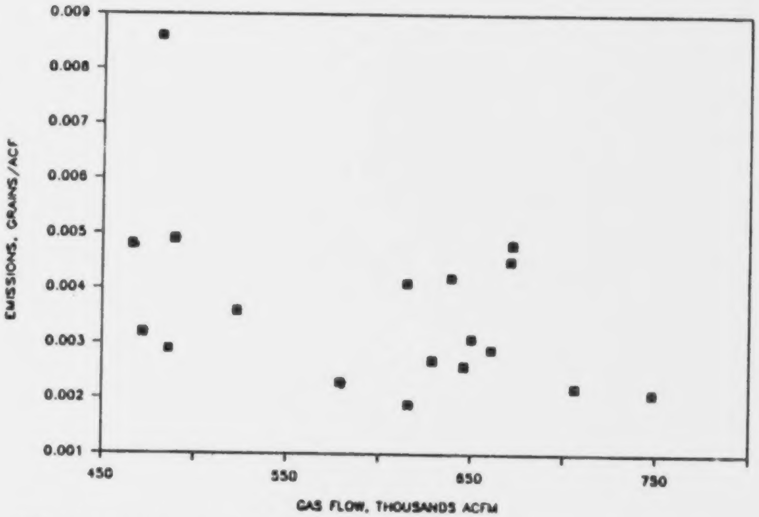

Figure 14. Emissions vs. gas flow, Parish.

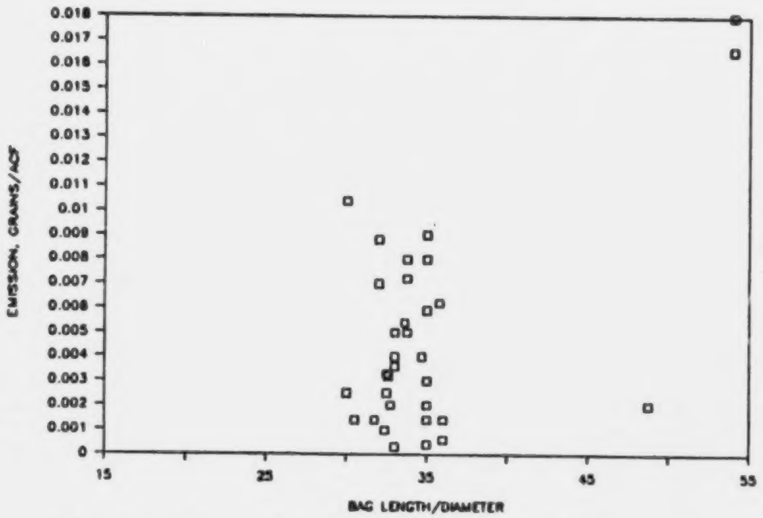

Figure 15. Emissions vs. bag L/D.

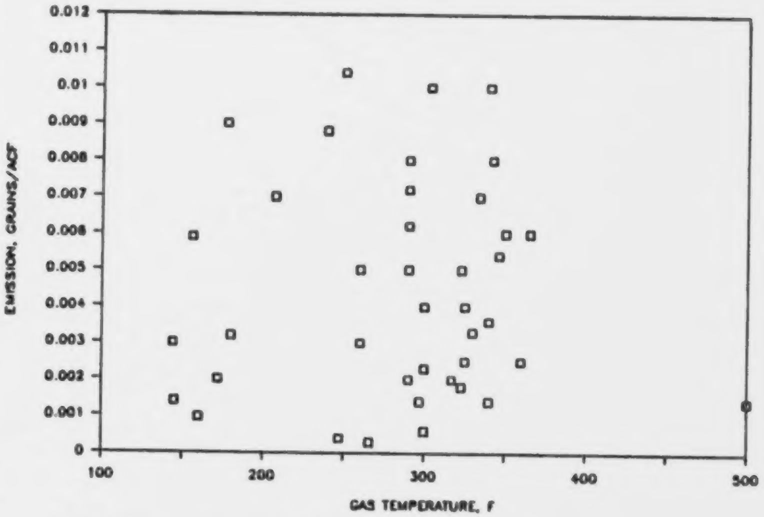

Figure 16. Emissions vs. temperature.

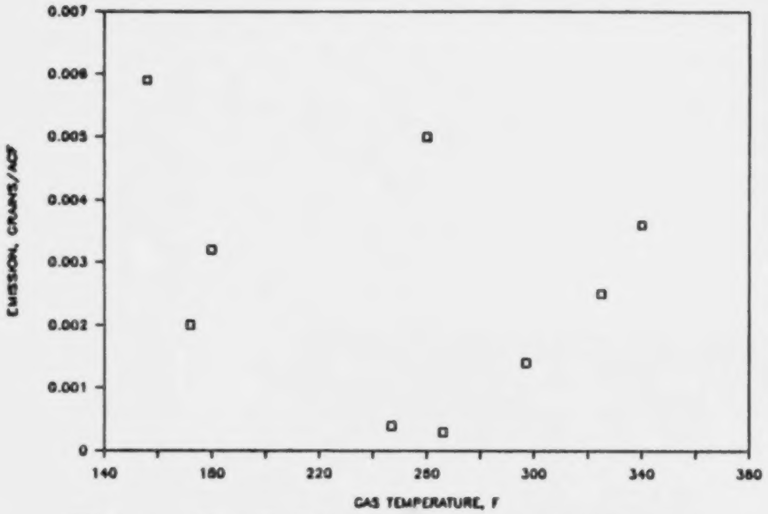

Figure 17. Emissions vs. temperature, one brand of fabric filter.

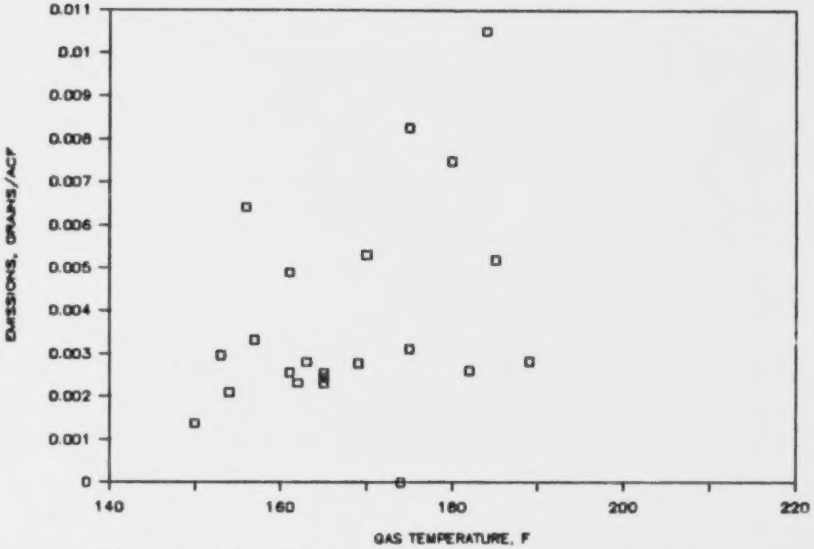

Figure 18. Emissions vs. temperature, pilot test results.

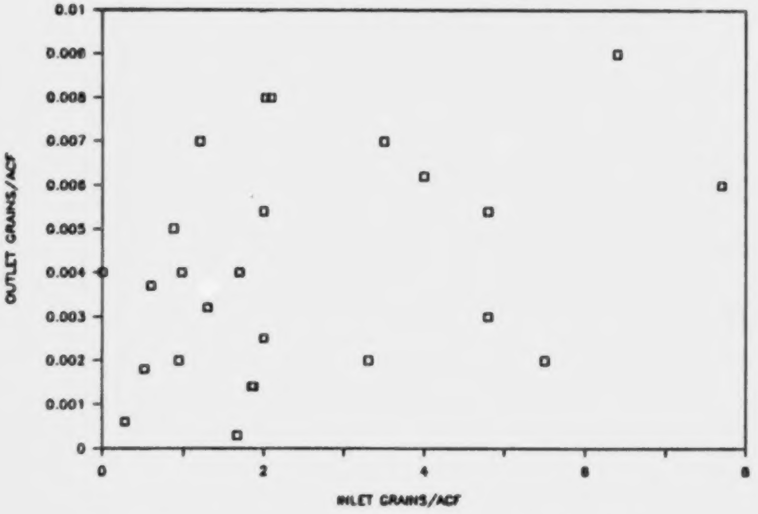

Figure 19. Emissions vs. inlet ash.

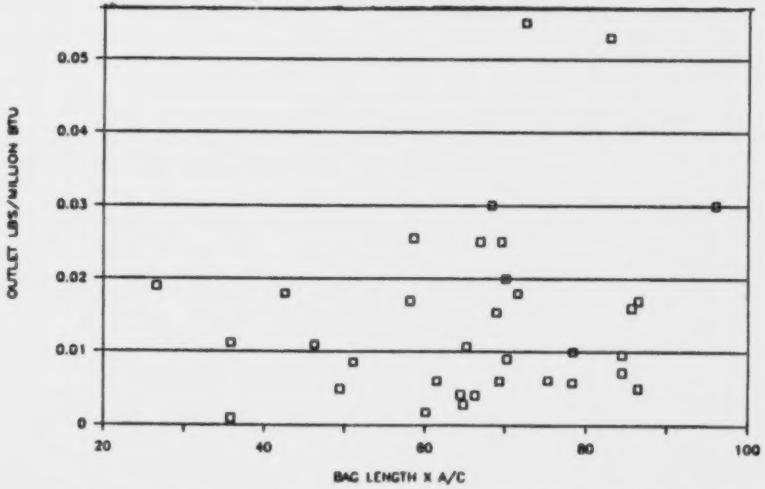

Figure 20. Emission vs. bag length X A/C.

TABLE 1

FABRIC FILTER EMISSION DATA

PLANT & UNIT	GRAINS/ACF	LBS/MM BTU	CLEAN	REF.
HOLTWOOD 17	0.0060	0.0400	SD	10&32
SUNBURY 1&2	0.0025	0.0060	R	42&32
NUCLA 2	0.0040	0.0080	SD	11,12&32
KRAMER 3	0.0006	0.0017	R	13&34
CAMEO 1	0.0020	0.0058	R	14&34
COORS 4	0.0100	0.0400	SD	15
ELK RIVER 3	0.0039	0.0170	R	16&34
SHAWNEE 5&6	0.0040	0.0160	R	17&33
HARRINGTON 2	0.0060	0.0200	SD	19&33
H. R. MILNER 1	0.0054	0.0156	PJ	36
MARTIN DRAKE 6	0.0014	0.0049	R	21&32
WAYNESBORO	0.0036	0.0110	R	22
CAT. TRACTOR	0.0038	0.0110	R	23
HANES	0.0037	0.0118	PJ	24
R. D. NIXON 1	0.0014	0.0041	R	25&37
ARAPAHOE 3	0.0003	0.0008	R	26
MISSOURI CITY 1&2	0.0081	0.0180	R	27&33
COORS 5	0.0079	0.0230	PJ	28
HARRINGTON 3	0.0060	0.0200	SD	29&34
BRUNNER ISLAND 1	0.0062	0.0180	R	32
ANTELOPE VALLEY 1	0.0020	0.0050	R	33
ANTELOPE VALLEY 2	0.0059	0.0170	R	33
ANTELOPE VALLEY 3	0.0020	0.0050	R	33
BULLOCK 1&2	0.0166	0.0250	R	32
CAT. TRACTOR	0.0023	0.0067	PJ	34
CHEROKEE 2	0.0072	0.0154	R	33
CHEROKEE 3	0.0050	0.0090	R	33
CLARK 1	0.0054	0.0190	R	34
CLAY BOSWELL	0.0070	0.0550	R	33
COYOTE 1	0.0070	0.0240	SD	32
CRAIG 3	0.0090	0.0200	R	33
CRISP 1	0.0104	0.0300	R	34
DULUTH 1	0.0014	0.0040	R	33
ESCALANTE 1	0.0020	0.0060	R	35
FOUR CORNERS 4&5	0.0088	0.0255	R	30
GEN. MOTORS	0.0052	0.0150	PJ	34
HOLCOMB 1	0.0032	0.0107	R	33
HUNTER 3	0.0030	0.0100	R	33
JAMES RIVER	0.0061	0.0180	PJ	41
KAW 1	0.0180	0.0530	R	33
KAW 3	0.0050	0.0086	R	33
LON D. WRIGHT 6	0.0033	0.0096	R	33
LON D. WRIGHT 7	0.0025	0.0072	R	33
LONGWORTH 3	0.0270	0.0957	PJ	32
NORTH VALMY 1	0.0010	0.0028	R	39
REID GARDNER 4	0.0080	0.0250	R	33
RIVERSIDE 4	0.0014	0.0040	R	40
RIVERSIDE 6	0.0004	0.0300	R	33
VALMONT 5	0.0023	0.0045	R	33
W.M. SAMIS 1	0.0200	0.0700	R	33
W.M. SAMIS 2	0.0300	0.0900	R	33
W.M. SAMIS 3	0.0100	0.0400	R	33
W.M. SAMIS 4	0.0200	0.0700	R	33
W.A. PARISH 8	0.0020	0.0061	R	6
WABAMUN 1	0.0200	0.0580	PJ	33
CROMBY 1	0.0120	0.0350	PJ	43
AVERAGE	0.0068	0.0218		

NOTES:

1. Where literature data shows only grains/acf or only pounds per million Btu, estimates have been made of the other units.

2. For some plants, various literature references may reflect more than one series of tests. For these plants, generally, the more current data are used.

FABRIC FILTRATION - STILL A BLACK ART

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ABSTRACT

Considering only the large custom-designed baghouses that are used to remove fly ash from the flue gas of coal-fired electric utility power plants, there are some things we do not know and some things we cannot explain. Our would-be scientific technology is still very much a black art.

This paper presents a different way to consider the variability of pressure loss components; discusses some anomalies in baghouse performance; offers some possible explanations for the anomalies; questions the correctness of our pressure loss formula; questions the way we use the formula; discusses the relative importance of cake weight and particle population and recommends some subjects for further research. In another paper (# 3B4) the author describes some new test procedures that will make it possible to do some research that we have not been able to do until recently.

PREDICTING PRESSURE LOSS

The literature includes many descriptions (1)(2) of baghouse start-ups where the pressure loss exceeded expectations. In some cases (3)(4)(5) side-by-side identical baghouses, cleaning the same flue gas, and with a short time between start-ups, have been reported to have significant differences in pressure loss. It is a fact that we do not know how to predict accurately the pressure loss for this application.

The formulas for predicting pressure loss are theoretical and each has its shortcomings. The one we use the most (6) is a combination of one that Darcy proposed in 1856 with a modification by Carman and Kozeny about 50 years ago.

Baghouse pressure loss has three components; the old cake loss, the new cake loss and the mechanical losses. The old cake (residual) loss is the pressure loss across the tubesheet immediately after a compartment has been returned to service after cleaning. The new cake loss is the pressure loss caused by the filter cake that accumulates on the bags in a compartment during the time it is on line between cleanings. The mechanical losses are the resistances to flow through the inlet and outlet ducts, through the inlet and outlet dampers, through the hopper and the clean side of the compartment. These three components vary for a compartment depending on where it is in the cleaning cycle but, the sum of the three always equals the flange-to-flange pressure loss for the baghouse. Our formula applies to the two cake losses. The mechanical losses are specific for each baghouse and cannot be predicted accurately; they can be estimated from a model study.

Our cake loss formula is:

$$\triangle P_t = k_1 V_t + k_2 C_i t V_t^n \quad \text{where,}$$

- $\triangle P_t$ is the pressure differential across the tubesheet and is the sum of the cake losses for one compartment at the end of time t, in.wg.,
- k_1 and k_2 are "constants" and conversion factors,
- V_t is the face velocity or, the velocity of the gas approaching and leaving the surface of the bag at time t, ft/min,
- C_i is the grain loading, gr/ACF,
- t is the time on line, min, and,
- n is an exponent usually assumed to be 2.

The first term, $k_1 V_t$ is the pressure loss, $\triangle P_r$, for the old cake. It includes the thimble entrance loss and the loss through the cloth, which are both negligible. We do not know k_1 or V_t but we do know that the product, $k_1 V_t$, is the pressure differential across the tubesheet immediately after the bags have been re-inflated and flow has been re-established after cleaning. The dimensions for k_1 are in.wg x min/ft.

The second term is the pressure loss for the new cake. The grain loading, C_i, is usually taken to be the grain loading entering the baghouse and is considered to be constant. It should be the grain loading entering the bags, which is different from that at the baghouse inlet due to hopper fallout and reverse air, RA, recirculation. It is not a constant because the velocity of the gas entering the bags decreases with time on line and the gas loses the ability to carry solids as that velocity decreases.

The time on line, t, is a variable that we control. Years ago we assumed that pressure loss can be reduced by reducing t, that is, by increasing cleaning frequency. Today we know that assumption is not always valid. The optimum cleaning frequency is also unpredictable.

The exponent for face velocity, n, is probably a number between one and two. We use the second power because that is what we use in other pressure loss formulas and because it is convenient. It is also possible that n may be greater than two.

The "constant", k_2, is a combination of several things. It serves as a conversion factor to convert weight of cake per square foot of bag area to in.wg. It also serves as a correction factor in case n is not two. Finally, and most important, it is an index of the permeability of the cake. Without k_2 the second term would give us weight per square foot. Pressure loss is proportional not only to weight per square foot; it is also proportional to cake permeability. We do not know how to predict k_2 nor have we determined it correctly for a baghouse in service. The dimensions for k_2 are in.wg x min x ft/lb.

WHAT GOES ON IN A COMPARTMENT

One reason for not being able to predict pressure loss is because we really do not know what goes on in a compartment. A baghouse is a complex, dynamic system with many things going on and all of them changing continuously. Our pressure loss formula uses average values for the variables. Average values do not provide a good representation for how those variables change during a cycle. We can approximate a better representation as follows.

The tubesheet pressure loss can be measured immediately after cleaning and again immediately before the next cleaning. The first measurement provides the old cake loss. The second provides the sum of the old and new cake losses. The new cake loss obtained by subtracting the old cake loss from the total of new and old is not correct because the old cake loss should be measured when t is as small as possible. At that time the face velocity, V, will be maximum. The sum of the old and the new cake losses, $\triangle P_{ts}$, should be measured when t is the time just before the next cleaning. The face velocity in the term for the new cake loss has to be the average face velocity during time t and that average V is not known. It varies with the compartment flow rate and with a compartment's position in the cleaning sequence. The grain loading, C_i, in the term for the new cake loss also has to be an average for the time on line. It too varies with flow rate and also varies with a compartment's position in the baghouse relative to the inlet duct because of RA recirculation. We do not know the average grain loading nor how much it is affected by RA recirculation.

The face velocity during time t can be calculated for a hypothetical case using a number of assumptions as follows. Assume values for k_1, k_2, and C_i. Assume that they are constants. Assume that n is two. Then assume that the flange-to-flange pressure loss, $\triangle P_{ff}$, immediately after cleaning the last compartment is 6 in.wg and that during 90 minutes with all compartments on line, that $\triangle P_{ff}$ increases to 8 in.wg. Also assume that the old cake loss is 2 in.wg immediately after the last compartment cleaned comes back on line.

If we were to monitor the pressure differential across the tubesheet of one compartment during the 90 minutes we would have the sum of the old and new cake losses from t=1 to t=90. If we subtract the tubesheet differential pressure, $\triangle P_{ts}$, from $\triangle P_{ff}$, we get the mechanical loss, $\triangle P_m$, for the 90 minutes.

If we then substitute our assumed values for k_1, k_2, and C_i and rearrange the formula we have:

$$\underbrace{(k_2 C_i t)}_{a} V_t^2 + \underbrace{k_1}_{b} V_t - \underbrace{\triangle P_{ts}}_{c} = 0$$

Solve for V_t by finding the positive root of,

$$V_t = \frac{-b \pm \sqrt{b^2 - 4ac}}{2a} \text{ for values of t from 1 to 90.}$$

Substituting V_t in the formula provides curves for the old and new cake losses. The results are shown graphically on Figure 1.

To obtain the curves for compartment No.1 we have to proceed in a different way. The pressure differential across No.1 immediately after cleaning will be approximately two inches greater than it was for No.20 and the flow rate through the compartment will therefore be greater. We can assume that the flow rate for No.1 will be $\sqrt{8/6}$ times the flow rate for No.20. In this case we cannot draw an arbitrary curve for $\triangle P_{ts}$ as we did for No.20. Instead we can increase $\triangle P_m$ as above and subtract that loss from $\triangle P_{ff}$ to get $\triangle P_{ts}$. No.1 differs from No.20 not only in flow rate it also differs in that $\triangle P_{ff}$ is decreasing instead of increasing for the first 60 minutes after cleaning and also that $\triangle P_r$ will be greater right after cleaning.

Making these changes and repeating the calculations we obtain the curves on Figure 2 which also shows the completion of the curves for compartment No.20.

These curves show that there are significant differences between compartment variables at the start and at the finish of a complete cycle. They also show differences between compartments dependent on place in the cleaning cycle. These results suggest that there may be some merit in changing the cleaning sequence by starting each sequence with a different compartment as one means to equalize compartment performance.

The literature includes numerous examples of the plot of fabric areal loading, W, versus filter drag, S, each complete with its own way of extending the linear part of the plot to obtain effective drag, S_E, and using the slope of the linear part of the plot to obtain k_2 which is equal to $\triangle S/\triangle W$. Some references (7)(8)(9) show a relatively long linear plot for new cake deposition and a relatively short non-linear plot when W has a low value that is not clearly defined. The upper end of this plot is the terminal drag, S_T, which is the drag just before cleaning and which can be determined by measuring the tubesheet pressure differential at that time and dividing it by some value of face velocity. The weight of cake at that time, W_T, is the sum of the old cake, W_R, plus the product of C_i x t x V where C_i and V have to be average values for the time between cleanings. To date we have not determined W_R directly. We can substitute S_E for k_1 in the term for old cake pressure loss provided we know both S_E and V_t when t is small. We do not know at which value of W the plot changes from curvilinear to linear. One reference (8) extrapolates the linear part of the plot to W_o which is where the drag is for the virgin fabric, and calls that intercept S_E. Another reference (9) extrapolates to the ordinate for W_R which is the residual (old) cake weight right after cleaning. The same reference increases the length of the curvilinear part of the plot and and decreases the length of the linear part. Remembering that these plots are for a single compartment it is reasonable to assume that the linear part of these plots probably corresponds to the linear portions of Figures 1 and 2 for V_t, $\triangle P_m$ and $\triangle P_r$.

These curves would be very different if we had reliable data for the variations of k_1, k_2, and C_i during a complete cycle. The curves in that case might help to explain some of our unknowns.

When we assume that k_1, k_2, and C_i are constants we are introducing error because they are all variables. Field and laboratory tests confirm that k_2 varies with face velocity. It also varies with gas viscosity, particle size, particle shape, particle bulk density, particle specific density, the ratio of particle surface area to particle volume, cake permeability and, in some cases to cake collapse. It also varies with time on line because cake composition and permeability change continuously during a cycle.

Grain loading, C_i, at the bag inlet will vary with time on line because the velocity entering the bottom of the bag will vary from as much as 900 ft/min to less than 100 ft/min in nominal 12 inch diameter bags. The rate of cake weight accumulation is thus variable and so are changes in particle size distribution. Both of these contribute to the variability of k_2.

If we use the cake loss formula to predict pressure loss by substituting average values for k_1, k_2, C_i, and V those averages will occur at different times for each variable in different compartments. The maximum P_{ff}, which is what we want to determine, will occur just before the start of the cleaning cycle and will be governed by the compartment that has been on line the longest. The average values for k_1, k_2, C_i, and V in that compartment at that time will be different from the values that will produce the cake losses and face velocities shown on Figure 2.

WHAT WE DO NOT KNOW ABOUT k_2

A lot. The Carman-Kozeny addition to Darcy's empirically derived formula is a theoretically derived term, k_2, which should help us to predict baghouse pressure loss. It has a formula:

$$k_2 = \frac{k \times \mu g}{g \;\; \delta_s} \times \frac{Ap}{\overline{V}p}^{\,2} \times \frac{1 - e}{e^3}$$

k = Carman-Kozeny constant, dimensionless

g = acceleration of gravity, ft/sec^2

μg = gas viscosity, lb(mass)/ft · sec

δb = bulk density of cake, lb/ft^3

δs = specific density of particle, lb/ft^3

Ap = surface area of particle, ft^2

$\overline{V}p$ = volume of particle, ft^3

e = porosity = void fraction = 1 - α

$\alpha = \frac{\delta b}{\delta s} = \frac{\text{bulk density of cake}}{\text{specific density of particle}}$, dimensionless

The absolute viscosity in the first term includes a pound mass dimension which is converted to pound weight by dividing by g. The specific density of fly ash can be determined from a specific gravity test. The only problem with the first term is k which varies with porosity. Carman and Kozeny recommended that their formula not be used for porosity greater than 0.7. Figure 3 shows the Kozeny relation between k and e for an assemblage of spheres.

The second term, specific area, is for solid spherical particles of a single diameter. The term, for spheres, reduces to $36/d^2$ which means that the term, and k_2, increase as particle diameter, d, decreases.

The third term, porosity, has not been determined exactly because that requires a measurement of the bulk density of the filter cake in place and in service at operating conditions.

Reference (10) reports the results of tests that confirm the occurrence of cake collapse. We can assume that cake permeability changes continuously during all the time a compartment is in service because the composition (particle size distribution) and thickness change continuously and therefore k_2 changes too. If, in addition, the cake should collapse at any time it would have a higher bulk density, greater k_2 and pressure loss. We do not know the conditions that cause cake collapse and we do not know how much difference it makes when it occurs. Finally, we do not know how to prevent it or to minimize it.

Darcy experimented with beds of materials in which the particles were all solids and had a variety of shapes and densities. Carman and Kozeny theorized about beds in which the particles were all solid spheres of the same diameter and density. The Darcy formula for flow and the Carman-Kozeny formula for k_2 are thus based on filters consisting of solid particles with no ability to absorb liquids and in which the voids consist of the flow paths which are the interstices between the particles. When we apply these formulas to fly ash we are using solid particle formulas for a material that includes many hollow spherical particles.

When we examine SEM micrographs of fly ash it is apparent; (a) that the predominant shape of the particles is spherical, (b) that broken spheres reveal that they are hollow with relatively thin shells, (c) that the shells include smaller spheres or voids, (d) that the larger spheres contain smaller spheres, suggesting that the smaller spheres probably contain even smaller spheres and that many of the non-spherical particles seem to be pieces of broken shells. We know that settling ponds used to separate fly ash from water do not work well because a large amount of the fly ash floats. We also know that fly ash does not absorb water. We can therefore assume that the spherical particle shells are not permeable. It thus seems reasonable to use the solid particle formulas even though most of the fly ash particles seem to be hollow.

Darcy's formula applies to fluids, both liquid and gaseous. For liquid flow through a filter bed we are familiar with the principle that resistance to flow is proportional to the surface area of the particles in the bed that are wetted by the liquid. The same principle applies to gaseous flow through a dry filter bed.

The reason is apparent if we use the hydraulic analogy. A single large diameter pipe with a certain flow rate will have a friction loss related to its internal circumference, the wetted surface. An array of smaller diameter pipes with the same total cross sectional area and flow rate as the big pipe will have a larger total of wetted surface and therefore a higher friction loss. A filter cake comprised of nothing but very small particles will have a high porosity relative to a cake of the same thickness but with larger particles. The pores in the first will be more numerous, with smaller cross sectional area, longer and more tortuous flow paths than in the second. The "wetted surface" in the first will be greater than in the second. The pressure loss through the first will be greater than through the second.

Although porosity increases as particle size decreases, suggesting that pressure loss will increase as particle size decreases, the Carman-Kozeny porosity term, $(1 - e)/e^3$, decreases as particle size decreases which means that k_2 and pressure loss will decrease as particle size decreases. The porosity term is thus in opposition to the specific area term. In order to determine the net result for the product of the two terms we need a relation between porosity and particle size for fly ash. The author, in a small search, was unable to find a correlation and elected to use data for powdered gypsum from reference (6). The correlation is shown on Figure 4.

Using Figure 4 to determine the porosity term for small particles and multiplying that by the specific area term we obtain an approximation of the net result for the product as shown on Figure 5. Note, from this figure, that a decrease in particle diameter from 10 microns to 1 micron results in a sixteen-fold increase in k_2 and pressure loss.

The literature contains a variety of data for fly ash specific density and for its bulk density. The author's assessment of some of these data is that fly ash specific density is reasonably constant regardless of the kind of coal and regardless of the type of combustion. A good value, for the purpose of this paper, is 160 lb/ft^3. Fly ash bulk density is apparently quite variable. Since we do not have data for it in place on bags in service we have to compromise with laboratory test data.

In addition to our inability to measure fly ash bulk density in the field we also have what seems to be the impossibility of measuring it in the laboratory. When we pass a measured flow of gas through a fly ash filter cake in the laboratory and calculate the porosity we are measuring only the void volume of the interstices between the particles. We do not, and probably cannot, determine the volume of the voids inside the unbroken hollow spheres of fly ash. We are not measuring the total void volume in that sample cake. If we could determine the total volume of the interstices plus that of the hollow spheres e would be larger than we currently think and as e increases both bulk density and pressure loss decrease. Using laboratory test data to determine porosity probably causes us to arrive at values for k_2 that are higher than they should be.

POPULATION OF SMALL PARTICLES

The foregoing references to small particles and to very small particles reflect the author's concern that we may have underestimated the smallness of the fly ash particles that we are trying to collect and also that the population of the small particles may be as significant, and possibly more significant, than the weight of the particles.

Our pressure loss formula, the collection efficiency and the allowable particulate emission limits are all based on weight. On a weight basis baghouses perform very well; on a particle population basis not well at all. That opinion derives from some estimates of fly ash particle populations that illustrate some additional things that we do not know and that we should be concerned about.

If we ask how many particles are there in a grain of fly ash we are posing a quodlibet; a question that has no answer. Perhaps the most famous quodlibet is the anonymous 600 year old question, "How many angels can dance on the head of a pin?" We cannot answer ours but we can make a crude estimate.

A grain of fly ash is a very small amount; it takes 7000 grains to make a pound. For a fly ash with a specific density of 160 lb/ft^3 a solid sphere of fly ash weighing one grain would be about the size of a peppercorn.

If we use that density and typical particle size distribution data we can calculate the number of solid spherical particles per size range using average particle sizes for the corresponding percentage ranges. If we do this for typical eastern and western bituminous coal fly ashes we obtain, in very round numbers, 4 billion particles per grain for eastern high sulfur coal and 144 billion for low sulfur western coal. We know that a very large percentage of fly ash particles are hollow spheres, cenospheres. Judging from some typical SEM micrographs it looks as though the shell thickness of the cenospheres is about 3 percent of the outside diameter. If we change from solid spheres to cenospheres with that shell thickness our populations increase about six-fold to 24 billion per grain for eastern coal and to 864 billion per grain for western coal.

If we do this calculation for a fly ash where we have BET (Brunauer, Emmett, Teller) data for the specific area term in addition to particle size distribution and specific gravity we can sum the surface areas and volumes for each of the size range populations to compare with the BET data. The calculated specific area term turns out to be a little smaller than the BET value. To make the calculated value agree with the BET value we have to increase the calculated value by about 15 percent. This increase may be explained by the fact that broken cenospheres are seen to contain many smaller particles when the size distribution was determined. On the other hand, the BET value may be too high for this purpose because it includes the interior surface of broken cenospheres and that surface probably does not affect flow resistance.

In any event, our crude estimate of particle population is then about 30 billion particles per grain for eastern coal fly ash and about one trillion for western coal fly ash. For both kinds of fly ash over 99 percent of the populations are particles smaller than the MMD. For the fly ashes used in these calculations the MMDs were 13 microns for eastern and 3 microns for western coal fly ashes.

When the electric utilities started to burn low sulfur western coal there were some unanticipated problems with particulate removal equipment. Problems with electrostatic precipitators were blamed on high fly ash resistivity. Some baghouses worked well and others did not. The problems of those that did not have not been explained. It is possible that some of the problems with the precipitators and most of the problems with the baghouses were due to a very significant difference in the number and size of small particles between eastern and western coal fly ashes.

We now have a new kind of particulate to contend with. That is the mixture of fly ash and chemicals from fluidized bed boilers. The large particles are removed from this mixture by cyclones ahead of the dust collector and recirculated to the furnace. The dust collector will thus receive a particulate with a large population of small particles. As in the case when shifting from eastern to western coal we may experience some more unanticipated problems when we apply dust collectors to fluidized bed boilers and for the same reason; large populations of very small particles.

SEM micrographs of fly ash raise several interesting questions. They show broken spheres containing many smaller unbroken spheres. If one kind of fly ash has sturdy shells that go from the furnace exit to the chimney without breaking we would have an eastern coal fly ash size distribution. If the shells are fragile and break easily somewhere between the furnace and the dust collector the small spheres contained in the large spheres would be released to produce a western coal type of fly ash size distribution.

So the first question is, do the differences in shell strength account for the differences in fly ash particle populations between eastern and western coals? Second, since almost all of the small spheres are unbroken are they hollow or solid? If they are hollow do they contain even smaller particles? Third, is there a ratio of shell thickness to particle diameter that separates those that break easily from those that do not?

We are proud of the fact that baghouses are inherently very efficient dust collectors on a weight basis. On a particle population basis we have a spur to do better. Consider a medium sized electric utility power plant burning a low sulfur western coal with a flue gas flow rate of a million ACFM and a baghouse particulate emission rate of 0.005 grains/ACF. Assume that there are a trillion particles per grain of fly ash and that they are all equal to or smaller than 3 microns. We would then have 5 quadrillion (5×10^{15}) particles per minute going up the chimney;

all very small particles. Billions per minute will be small enough to serve as condensation nuclei for the oxides of sulfur and nitrogen going out the same chimney and thus helping to form acid rain.

There has been very little improvement in baghouse design and very little, if any, reduction in baghouse particulate emission rates since the start-up of the Sunbury baghouses in 1973. The emphasis during these past fifteen years has been on trying to improve an old design. That kind of research and development is not good enough. We need a new approach; something other than remedies for old ills. We can expect that the PM-10 ambient standards will lead to similar particulate emission regulations in the future. If we do not start doing some new kinds of research soon we will not be ready when that whistle blows.

The intent of this paper is to serve as a reminder that there are a lot of things we do not know about fabric filtration and to stimulate interest in new efforts to acquire some new knowledge. We have a long way to go towards being a scientific technology. Our present state-of-the-art is still a black art.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

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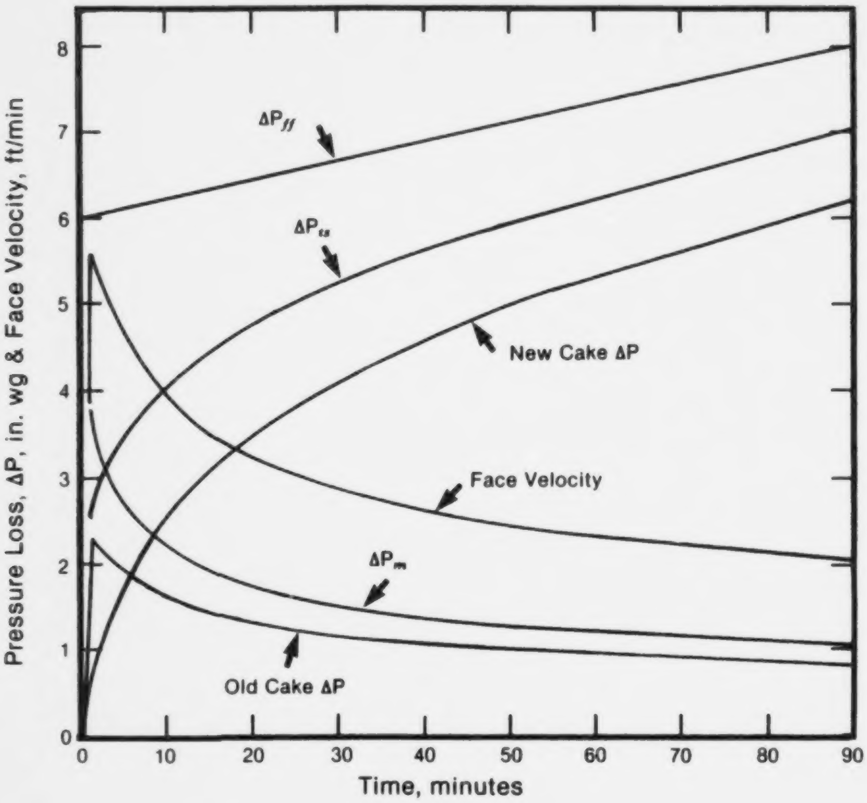

Figure 1. Variables for Compartment 20

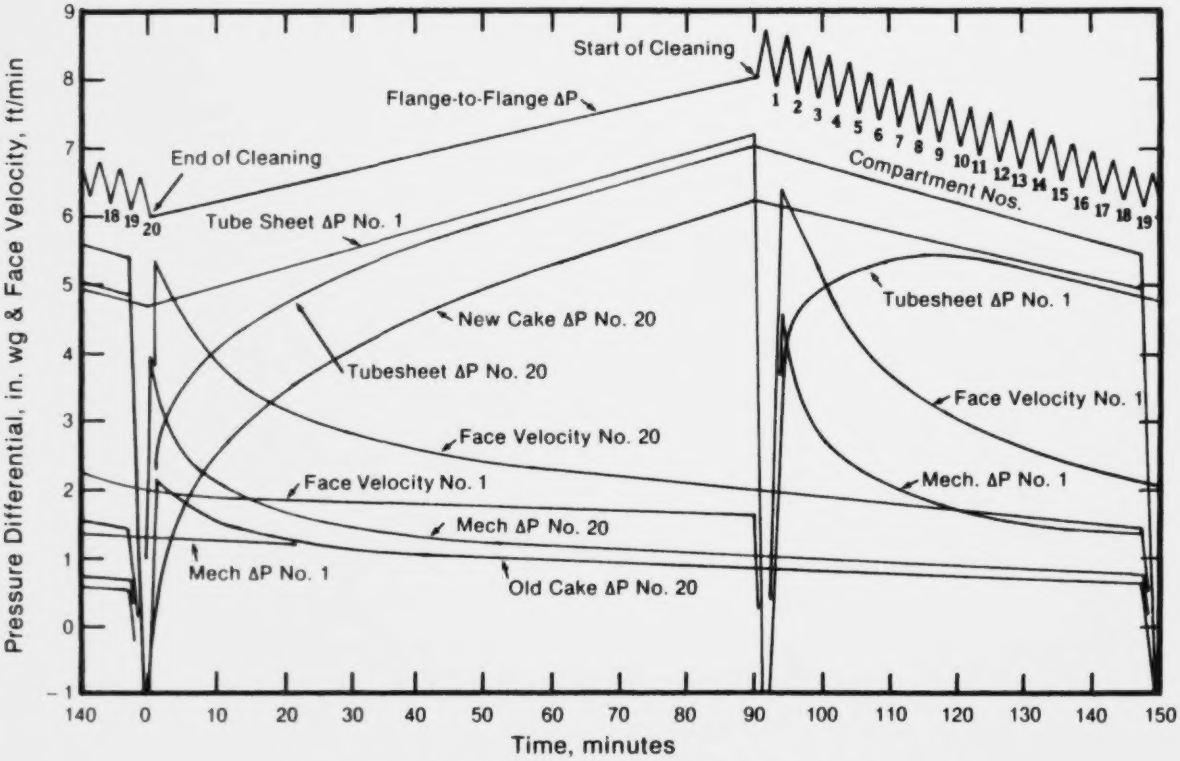

Figure 2. Variables for Compartments 1 & 20

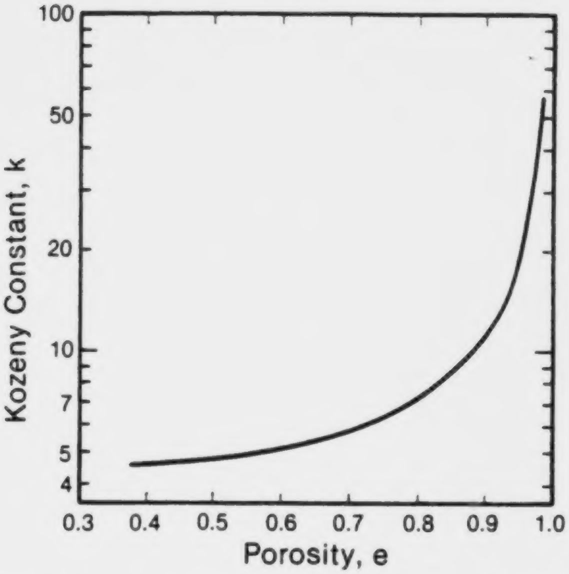

Figure 3. Kozeny Constant vs. Porosity (6)

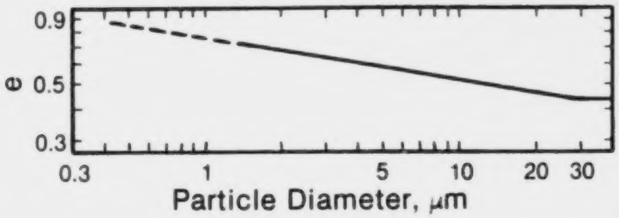

Figure 4. Porosity vs. Particle Diameter (6)

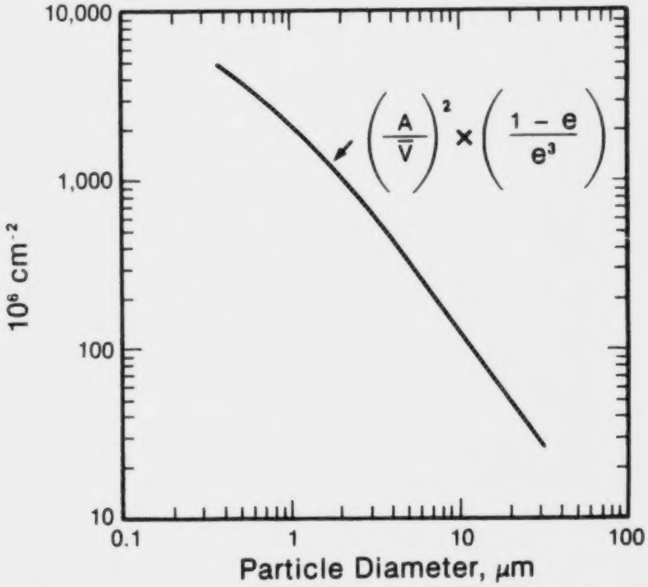

Figure 5. (Specific Area)(Porosity Term) vs. Particle Diameter

Flue Gas Dust Extraction for the Slag-tap Pulverized Coal Firing At Brunswick Central Thermal Power Station, Using the "Superjet" Tubular Filter

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This paper reports on the advantages of TEFAIRE with regard to the investment and operating costs of flue gas purification plant compared with TEFLON needle felts for tubular filter installations.

FIGURE NO. 1

BUHLER BROTHERS Ltd. Head Office and production plant in Uzwil/-Switzerland.

Please let me first give some information about our company worldwide activities:
The BUHLER group of companies consists of BUHLER BROTHERS Ltd. Uzwil/Switzerland with 5 branch factories in Switzerland with
. 3500 employees
BUHLER-MIAG Ltd. Braunschweig/Germany with
. 1000 employees
25 Bühler-Miag companies and offices and 80 agencies around the world with
. 2000 employees
in total about
. 6500 employees.

Consolidated turnover per year over 1 billion SFR. Bühler is known worldwide in the flour milling and food industry for its famous grain mills; grain elevators; macaroni factories; food and feed processing plants.
The nonfood sections comprise processing of plastics, rubber and various other products; metal diecasting equipment; ship unloading and loading systems; mechanical and pneumatic continuous conveying systems; environmental protection plants such as recycling or disposal of refuse and sewage sludge; and last but not least our dust collection filter and fly ash handling systems. For more information about BUHLER Group of companies please see our booklet "One Group - Different Markets" and our Manufacturing Program.

SUMMARY

BUHLER SUPERJET^R baghouse with Du Pont **TEFAIRE**^{R} needle felt filter bags.

FIGURE NO. 2

160 MW slag-tap pulverized coal fired Thermal Power Station at Central Brunswick, West Germany, equipped with **BUHLER SUPERJET**^{R} baghouses using **DU PONT TEFAIRE**^{R} filter bags.

BUHLER flue gas filtration systems by means of **BUHLER-SUPERJET**^{R} baghouses using **DU PONT TEFAIRE**^{R} filter bags ensure high efficiency and low downtimes for coal-fired industrial boilers, such as fluid-bed type, stoker-type or slag-tap pulverized coal-type.

FIGURE NO. 3

BUHLER SUPERJECT^R baghouse with 2668 **TEFAIRE**^{R} filter bags at the Brunswick plant.

BUHLER-SUPERJET^R baghouses are equipped with a patented high-performance direct-pulse bag cleaning device, and with **TEFAIRE**^{R} needle-felt filter bags with high gas permeability. The combination of this two new elements makes it possible to run this type of flue gas filtration system at a higher gas to cloth ratio with controlled differential pressure. This is a big advantage compared with traditional reverse-flow baghouses using woven fabric **TEFLON**^{R} bags or the conventional venturi pipe reverse jet filters with needle felt **TEFLON**^{R} bags. The following table shows gas-to-cloth ratios, as they are typical for the different filter types and bag materials:

TABLE NO. 1

Gas-to-cloth ratio typical for different filter cleaning devices.

Type of Filter equipment	Traditional Baghouses	Conventional Baghouses	**BUHLER SUPERJET**^{R} Baghouses
Cleaning material	Reverse flow	Venturipipe reverse jet	Direct pulse
Filter media	Woven **TEFLON**^{R} bags	Needle felt **TEFLON**^{R} bags	Needle felt **TEFAIRE**^{R} bags
Gas-to-cloth ratio $m^3/m^2 \cdot min$ (m/min)	0,5	1,0	1,5
Gas-to-cloth ratio ft/min	1,62	3,24	4,86

Flue gas filters operated at above gas-to-cloth ratios will run with a pressure differential across the filter bags of about 150 - 200 mm WC (approx. 6 - 8 inches of water) and with less than 2-5 mg/m^3 (less than 0,010 grains/cu. ft.) residual dust content in the cleaned gas. Pressure difference and residual dust content are depending on quality of coal and type of boiler. Same filters can be operated without desulphurisation of flue gas to collect the fly ash, or with spray dryer desulphurisation to collect calcium sulphates. The combination of BUHLER and DU PONT technology makes it possible to design and to install smaller filter units, which need less volume. Compared with traditional baghouses, smaller buildings are possible. This new type of filtration equipment is more competitive and easier to maintain.

FIGURE **NO.** 4

Clean gas side of filter opened for bag inspection.

All maintenance is possible from the outside. No Personnel does not have to enter into dirty aereas inside of hot flue gas chambers. Separation efficiency is independant to various conditions of flue gases, independant to the dust concentration and independant to sulphur dioxyd, humidity and temperature. All these facts makes **BUHLER SUPERJET**^{R} filters with **TEFAIRE**^{R} bags competitive to traditional and conventinal baghouses as well as to electrostatic precipitators.

TABLE **NO.** 2

Pressure differences as a function of gas-to-cloth ratios for different needle felt medias at 30°C (67°F).

TABLE **NO.** 3

Pressure differences as a function of gas-to-cloth ratios for different needle felt filter medias at 140°C (310°F).

Results from a special test-rig which is installed in the BUHLER-owned power station, showing pressure differentials and residual dust contents as a function of gas-to-cloth ratios for different needle felt filter media such as **TEFLON**^{R} 9456, Glas 850, **TEFAIRE**^{R} 800 and **TEFAIRE**^{R} 500.

TABLE **NO.** 4

Residual dust content as a function of gas-to-cloth ratios for different needle felt medias 30°C (67°F).

The extreme triboelectrical characteristics between glass fibers and TEFLON^{R}-fibers must be responsible for the extremly good results of **TEFAIRE**^{R} 500.

Good experiences since three years, in different coal fired power stations, prooves that this new BUHLER/DU PONT filter technology can be a great help to electricity industry in finding better solutions fo their flue gas cleaning problems.

BUHLER CHAIN CONVEYORS for fly ash discharge and transportation.

FIGURE **NO.** 5

BUHLER CHAIN CONVEYORS for coal fired power stations and BUHLER-WIDE-PRECIPITATOR DUST DISCHARGER

Dusttight, abrasion-resistant, low-maintenance **BUHLER-CHAIN-CONVEYORS**, have prooven to be excellent technical and commercial solution for extracting fly ash from hoppers of any type of baghouse or electrostatic precipitators and for further transportation to fly ash bins or fly ash disposal areas.

FIGURE NO. 6

TUTUKA-Powerstation-Dusthandling plant general information.

Many of this fly ash conveyors have already been built by BUHLER. The latest coal-fired thermal power stations in South Africa were equipped with BUHLER fly ash chain conveyors for discharging electrostatic precipitators.
. the TUTUKA-plant with 6 x 600 MW-blocks
. and the MATIMBA plant with 6 x 650 MW-blocks
Explosion proof coal feeding systems and fly ash extraction and transportation systems where delivered by BUHLER-MIAG Sout Africa.

BUHLER-MIAG, Minneapolis, USA quoted through Black and Veatch fly ash extraction systems, for Taiwan Power, Taichneg Power Station Project. Fly ash discharge of 32 hoppers below precipitator with ash silo holding 3520 tons.

BUHLER-SCRAPERS

Storage and transfer plants for calcium sulfates (Gypsum), which is a byproduct from desulphurization of flue gases, are equipped with BUHLER scrapers and chain conveyors.

BUHLER PNEUMATIC CONVEYING

To complete the manufacturing program, BUHLER, specialized in pneumatic conveying. Pioneering work has also been done in developing the technology of pneumatic low-velocity conveying. Between 1980 - 1984, the US Vaaler Awards was given three times to BUHLER-MIAG INC. Minneapolis, for introducing this new technology to the market.

Figure No. 7

US Vaaler Awards (Chemical Processing Nov. 1980 and 1982) Honors winner for **TAKT-SCHUB**^{R}.; pneumatic low velocity conveyors.

Figure No. 8

US Vaaler Award (Chemical Processing Nov. 84) Top Honors for **FLUID-SCHUB**^{R}. pneumatic low velocity conveyors.

Figure No. 9

US Vaaler Award Honors wimmer for FLUIDSTAT^{R} pneumatic low velocity conveyors.

"Benefits fo BUHLER-MIAG low velocity conveying system.

Only about 60 % of the kW-requirement, and about 27 times less wear in pipes and elbows are typical for **BUHLER FLUIDSTAT**^{R} pneumatic low-velocity conveyors, compared with conventional high-density conveyors.

The pulverised coal-fired Central Brunswick power plant West Germany, is equipped with a total of 11 **FLUIDSTAT**^{R} low-velocity pneumatic conveying systems for the handling of fly ash and calvium sulphate (Gypsum)

Figure **No.** 10

FLUIDSTAT^{R} pneumatic low-velocity conveyors for fly ash and calcium sulphate, at Central Brunswick Plant, West Germany.

Figure **No.** 11

BUHLER deliveries for flue gas cleaning baghouses, fly ash and gypsum conveyors, including the electrical controls.

BUHLER offers the design and supply of the dust collecting baghouses, fly ash and gypsum handling systems by one and the same company, which includes many advantages to consulting engineers and customers.

ELECTROSTATIC PREUPITATORS VERSUS FIBER FILTRATION

Table No. 5

Investment costs index as a function of residual dust content in the clean gas.

Separation efficiency of electrostatic precipitators depends very much to the variable composition (sulphur dioxide, humidity, temperature, dust concentration) of the flue gases and the specific electrical resistance of the dust itself, while separation efficiency of needle felt bag filters with its mechanical retention effects is very much independend to such variables.

Investment cost index as a function of residual dust content, for electrostatic precipitators compared with needle felt baghouses, shows the attractivity of the mechanical fiber filtration with lower emission limits.

If the investment costs index for a residual dust concentration of 150 mg/m^3 (0,0652 gr/cu. ft.) ist taken as 100, the investment costs index for a residual dust concentration of 5 mg/m^3 (0,011 gr/cu. ft.) rises up to 200 while costs for the baghouses remain more or less constant.

Table NO. 6

Arguments for fiber filtration versus electrostatic filtration.

REMARKS ON FIBERS AND LITERATURE

Earlier development of fiber filters where handicapped by the poor resistance of natural fibers to high temperatures and chemical agressions.

Since it is possible to produce suitable syntetic fibers the market of fiber filtration growed very fast. Applications for flue gas cleaning with Nomex, Rylon, Teflon are known since about 15 years. Since the new **TEFAIRE**^{R} already shows 3 years excellent results from industrial power plants, the fiber filtration becomes even more attractive.

Very interesting studies concerning investment and operating costs of electrostatic precipitators and different type of baghouses are concentrated in a paper of W.F. Schultz, given at the Du Pont Air Pollution Control Symposium 1987 at Basle, Switzerland.

The microscope photographes Figure No. 12 to 15 are taken from the paper, presented by W. Flatt at the same symposium, showing the **TEFAIRE**^{R} fiber configuration.

TEFAIRE^R is a mixture of about 15 % glass fibers and 85 % **TEFLON**^{R} fibers. Thickness of glass fibers is about 3-4 u, thickness of **TEFLON**^{R} fibers about 40 u (microns).

Figure No. 12

Photography by scaning electronic microscope of clean **TEFAIRE**^{R} needle felt 1000 times enlarged.

Figure No. 13

Photography by scaning electronic microscope of dusty **TEFAIRE**^{R} needle felt 1000 times enlarged.

Figure No. 14

Photography by polarised light microscope of clean **TEFAIRE**^{R} needle felt 40 times enlarged.

Figure No. 15

Photography by polarised light microscope of dusty **TEFAIRE**^{R} needle felt 40 times enlarged.

More detailed informations about the positive industrial experiences with **TEFAIRE**^{R} filter bags at the 160 MW power station, Central Brunswick, West Germany can be taken out from above mentioned paper.

FURTHER INFORMATIONS

So far my concentrated informations about our activities and experiences with new **BUHLER-SUPERJET**^{R} baghouses using **DU PONT TEFAIRE**^{R} filter bags for flue gas cleaning; with **BUHLER-CHAIN-CONVEYORS** for explosion and dust proof methering, feeding and transportation of coal wear resistant dust proof extraction and transportation of fly ash from electrostatic precipitators or baghouses, with BUHLER-SCRAPPERS for storage and transfer of gypsum, with BUHLER FLUIDSTAT pneumatic low velocity conveyors for extraction and transportation of fly ash, calcium hydroxide and calcium sulphate.

I am prepared to give more detailed informations and technical dates about coal fired 160 MW Brunswick plant, about the 22 MW and 80 MW Frankfurt plant and the 40 MW Karlsruhe plant. I have some more overhead transparencies of different other smaller flue gas cleaning **SUPERJET** bag filters.

I am also prepared to discuss more into details on the other subjects, by means of overhead transparencies.

It depends on your questions and primary interest to the different subjects.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therfore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

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   Brunswick Municipal Authority - Central Thermal Power Station.
   ETV-Register 41 (1982)

2. Klinger, R. and F. Löffler:
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   Chem. Eng. Process. 18 (1984), pp. 189-198

3. Löffler, Dietrich and Flatt:
   Dust precipitation with tubular and compartment-type filters.
   Publ. Friedr. Vieweg u. Sohn Verlag, Braunschweig/Wiesbaden.

4. Husistein, H. and W. Peters:
   Flue gas dust extraction plant for the Brunswick Central thermal power station.
   Bühler "Diagramm" 82

5. H. Michele:
   Flue gas purification with dry absorbents - potential and limitations.
   Chem. Ing. Tech. 56 (1984) No. 11, pp. 819-829

Figure No. 1 **BUHLER BROTHERS Ltd.**
Head Office and production plant

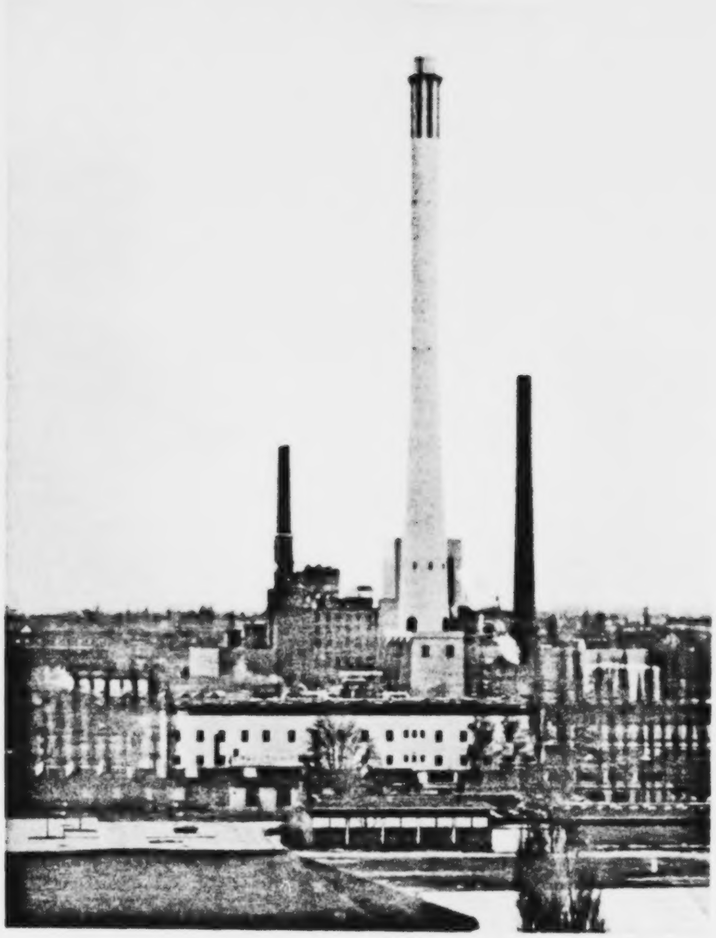


Figure No. 2 160 MW slag-tap pulverized coal fired Thermal Power Station at Central Brunnswick West Germany.

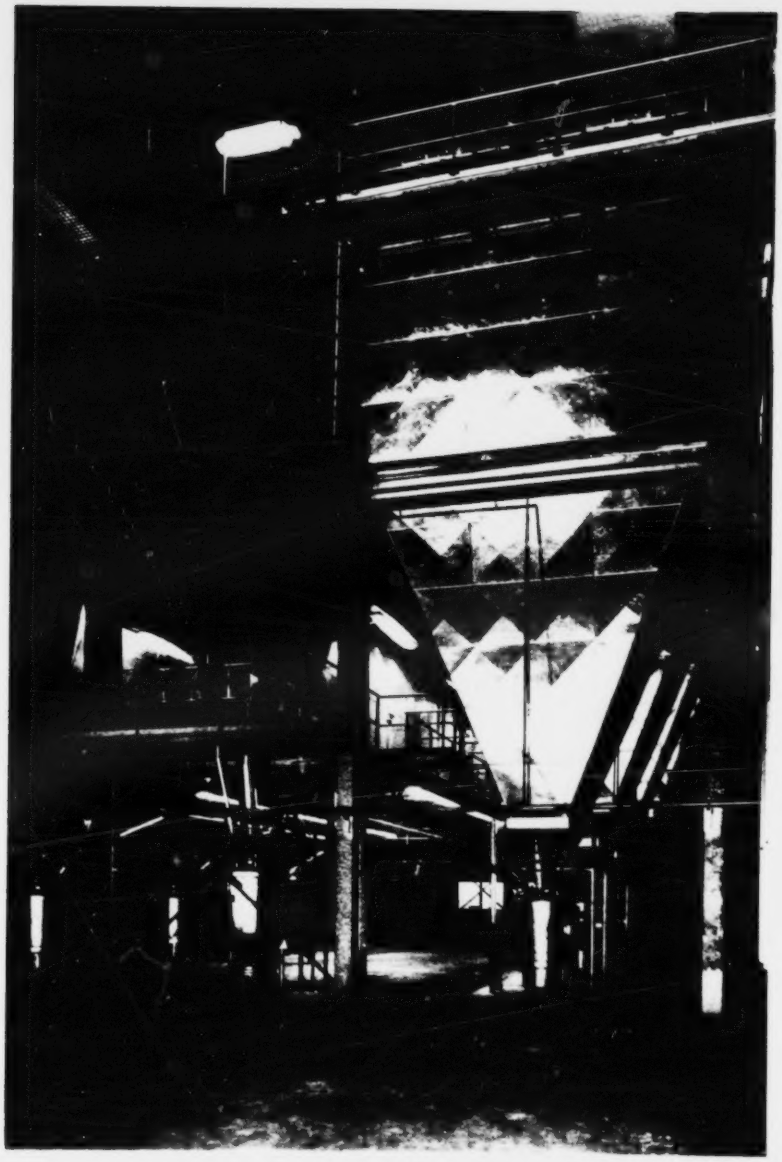


Figure No. 3 BUHLER SUPERJET^R baghouse with 2668 TEFAIRE^R filter bags at the Brunswick plant.

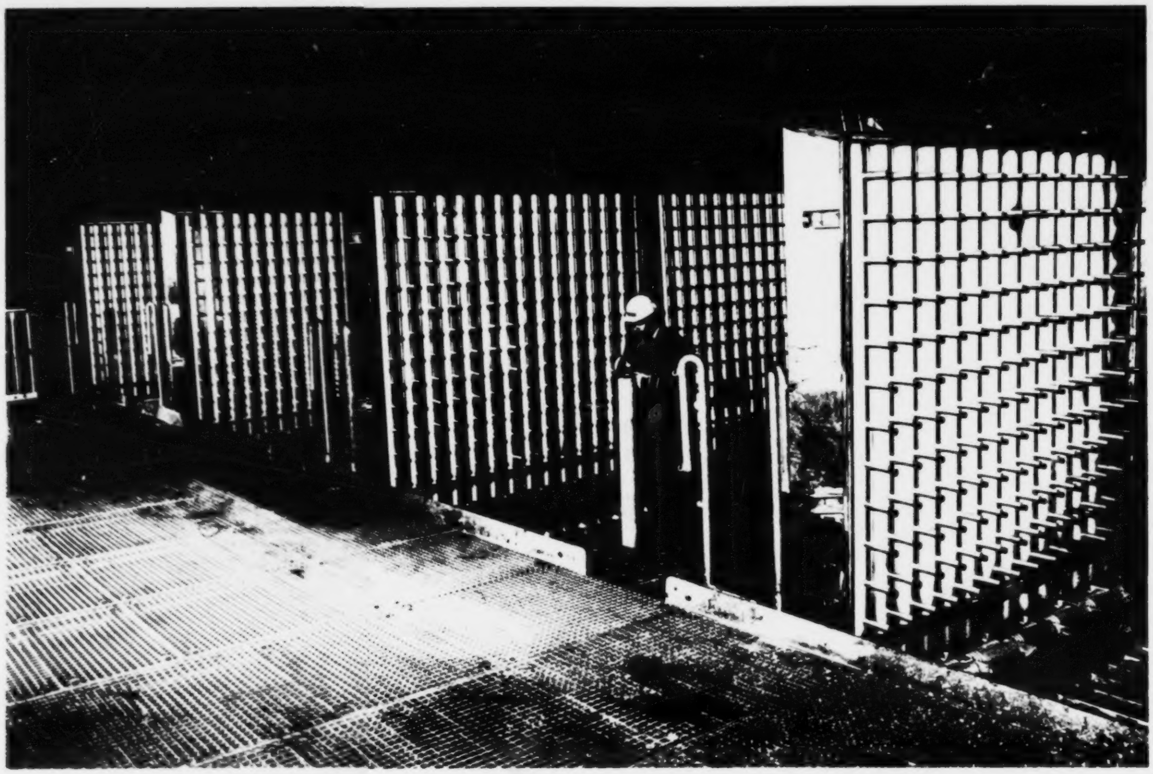
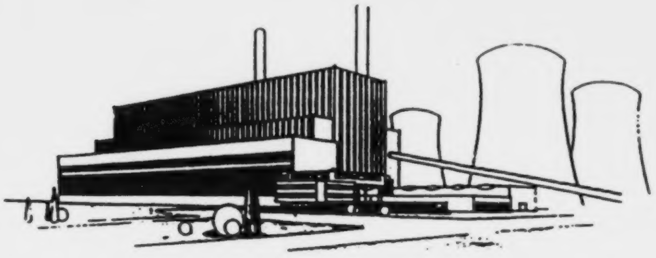


Figure No. 4 Clean gas side of filter opened for bag-inspection.

BUHLER CHAIN CONVEYORS FOR COAL-FIRED POWER STATIONS



E.D.F. Power Station, Bouchain, France

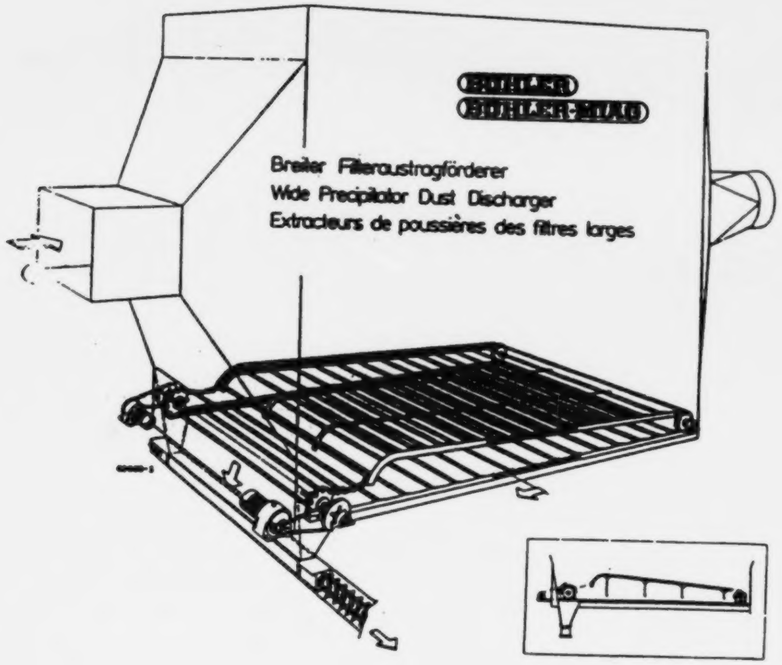

BUHLER
BUHLER-MIAG
Breiter Filteraustragförderer
Wide Precipitator Dust Discharger
Extracteurs de poussières des filtres larges

Figure No. 5 BUHLER cain conveyors for coal fired power stations and BUHLER wide precipitator dust discharger

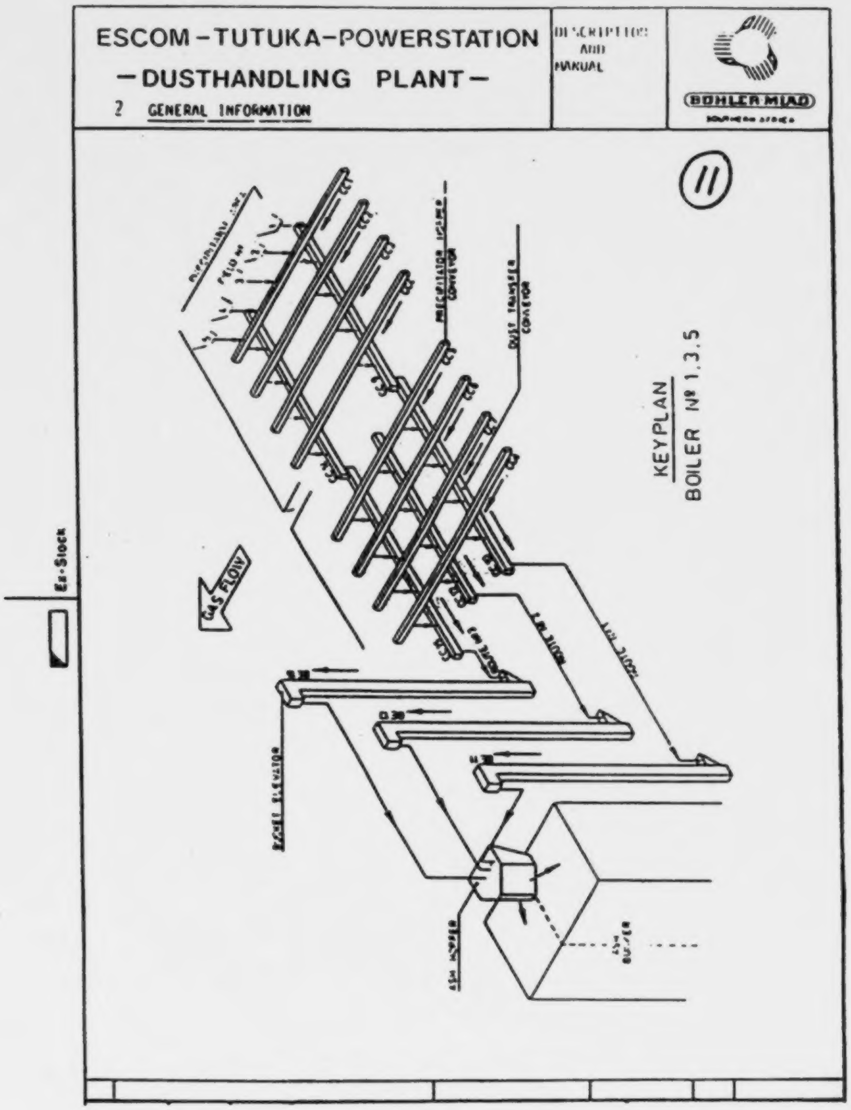
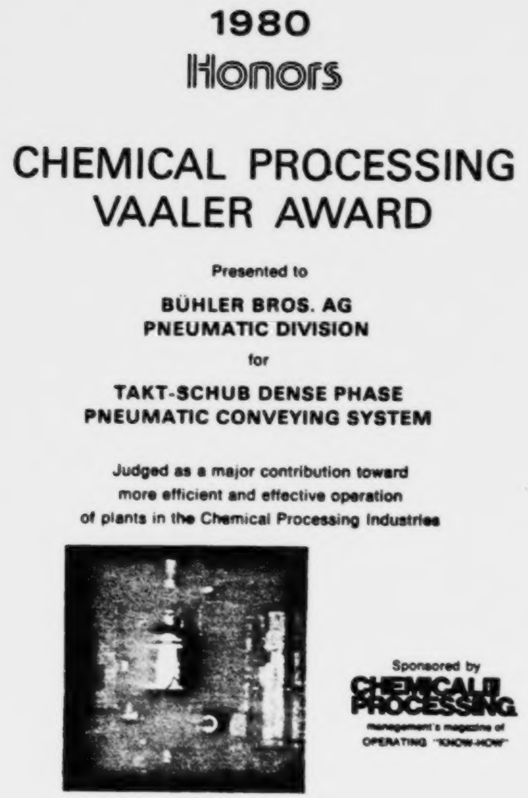


Figure No. 6 TUTUKA-Powerstation Dusthandling by BUHLER chain conveyors.



1980

Honors

CHEMICAL PROCESSING
VAALER AWARD

Presented to

BÜHLER BROS. AG
PNEUMATIC DIVISION

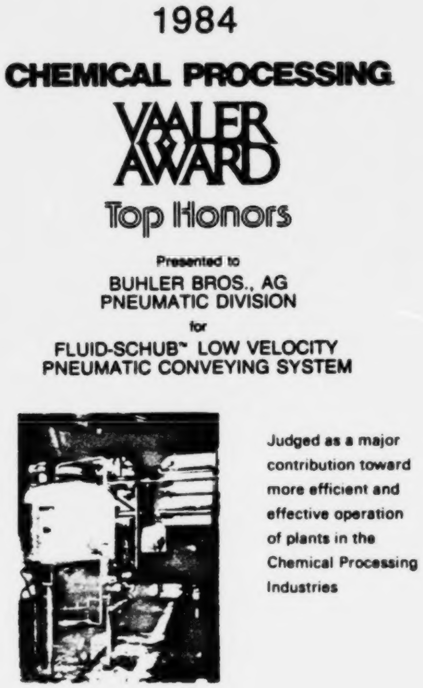
for

TAKT-SCHUB DENSE PHASE
PNEUMATIC CONVEYING SYSTEM

Judged as a major contribution toward
more efficient and effective operation
of plants in the Chemical Processing Industries

Sponsored by
CHEMICAL PROCESSING
management's magazine of
OPERATING "KNOW-HOW"

Figure No. 7 US Vaaler Award Honors winner for TAKT-SCHUB^R pneumatic low velocity conveyors.



1984

CHEMICAL PROCESSING

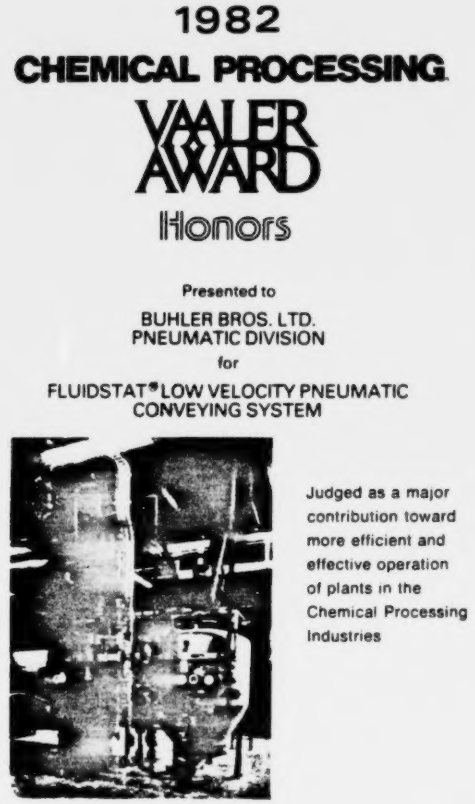
VAALER AWARD

Top Honors

Presented to
BUHLER BROS., AG
PNEUMATIC DIVISION
for
FLUID-SCHUB™ LOW VELOCITY
PNEUMATIC CONVEYING SYSTEM

Judged as a major
contribution toward
more efficient and
effective operation
of plants in the
Chemical Processing
Industries

Figure No. 8 US Vaaler Award Top Honors wimmer for FLUID-SCHUB^{R} pneumatic low velocity conveyors.



1982
CHEMICAL PROCESSING
VAALER AWARD
Honors

Presented to
BUHLER BROS. LTD.
PNEUMATIC DIVISION
for
FLUIDSTAT® LOW VELOCITY PNEUMATIC
CONVEYING SYSTEM

Judged as a major contribution toward more efficient and effective operation of plants in the Chemical Processing Industries

Figure No. 9 US Vaaler Award Honors winner for FLUIDSTAT^R pneumatic low velocity conveyors.

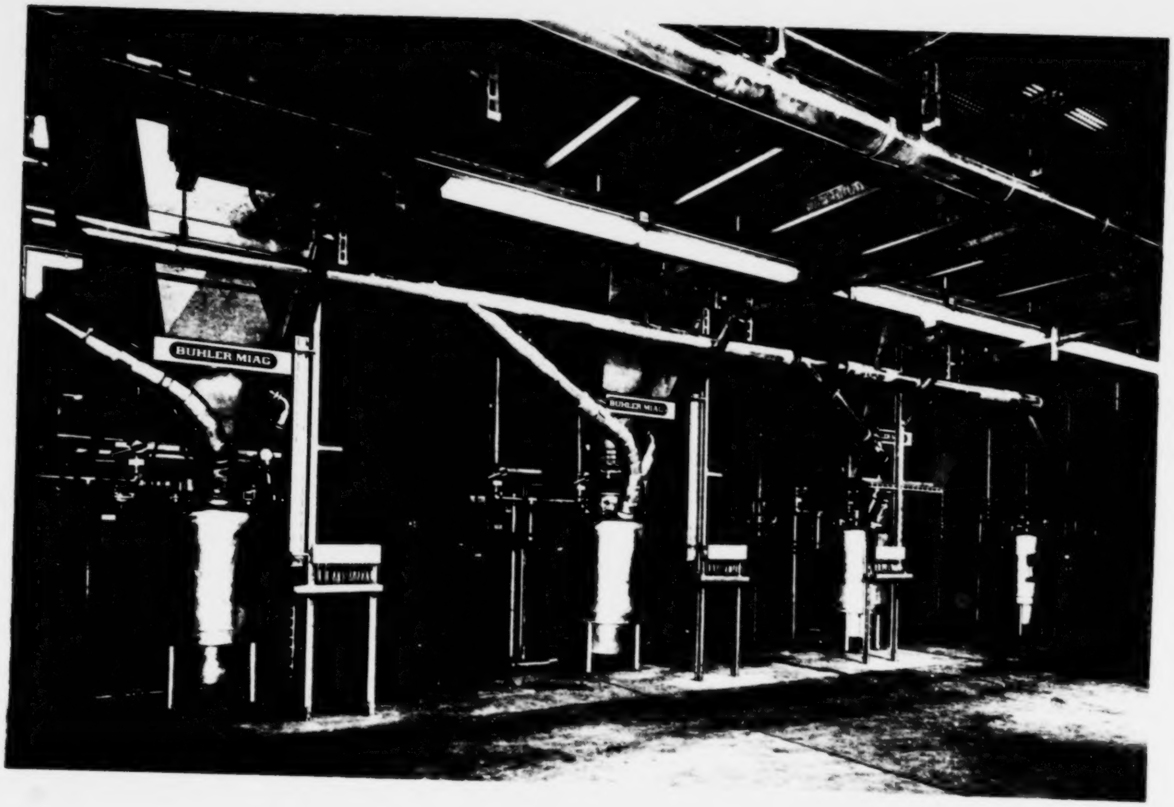


Figure No. 10 FLUIDSTAT^R pneumatic low velocity conveyors for fly ash and caltium sulphate, at 160 MW pulverised coal fired thermal Power station, Central Brunswick.

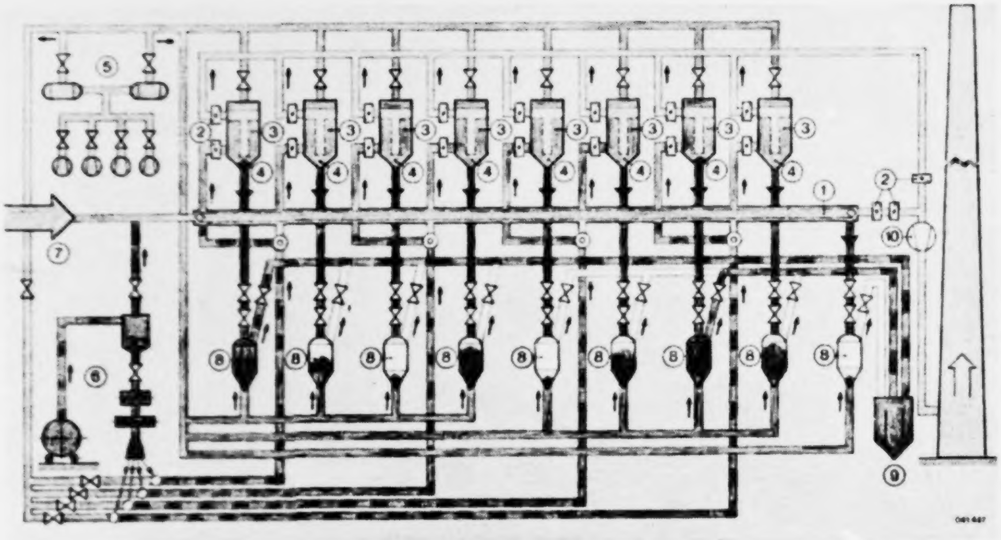


Figure No. 11 Control panel for BUHLER-SUPERJET^R baghous and FLUIDSTAT^R pneumatic low velocity conveyors at Central Brunswick plant.

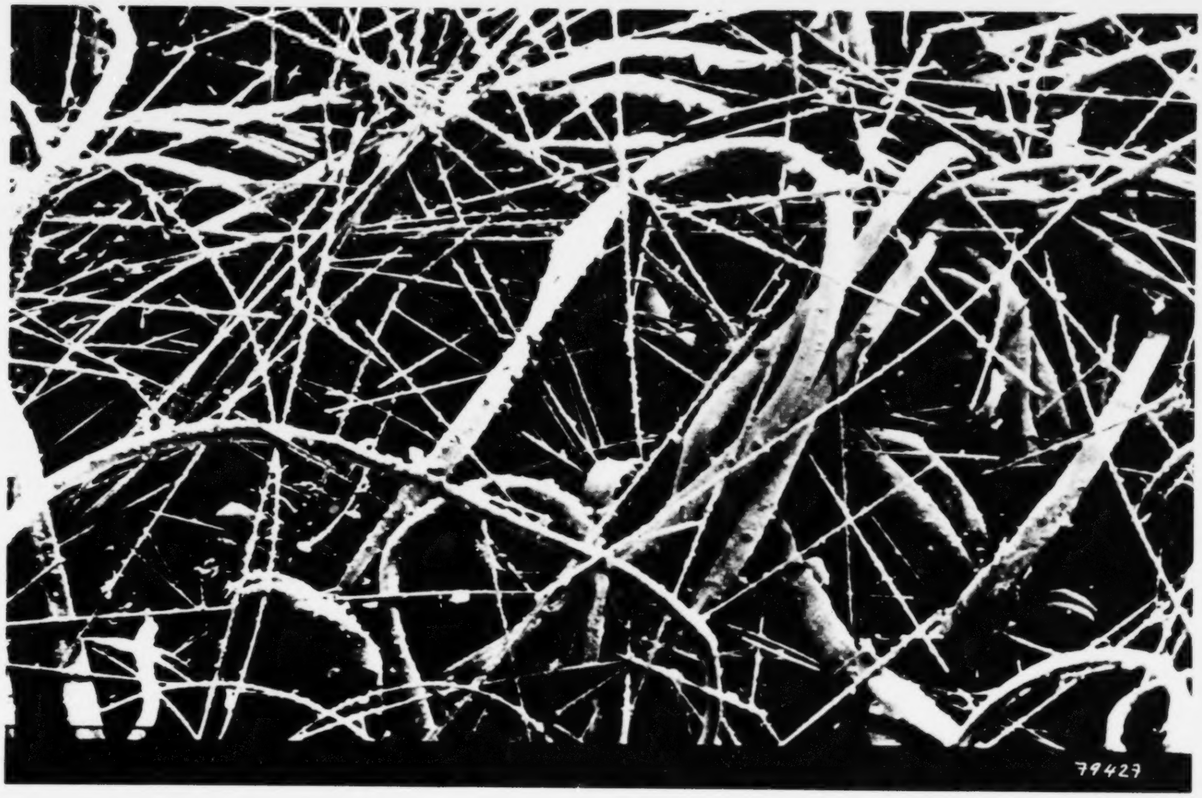


Figure No. 12 Photography by Scaning electronic microscope of clean TEFAIRE^R needle felt 1000 times enlarged.

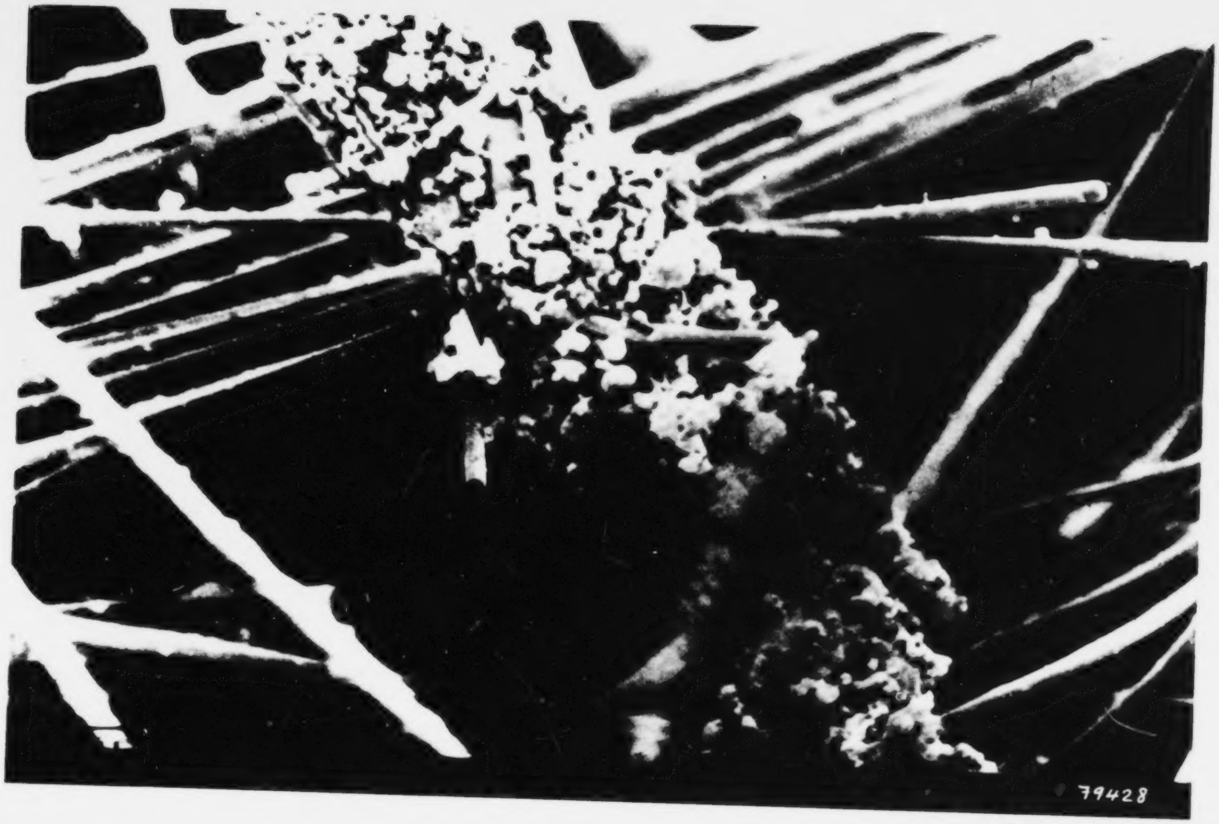


Figure No. 13 Photography by Scaning electronic microscope of dusty TEFAIRE^R needle felt 1000 times enlarged

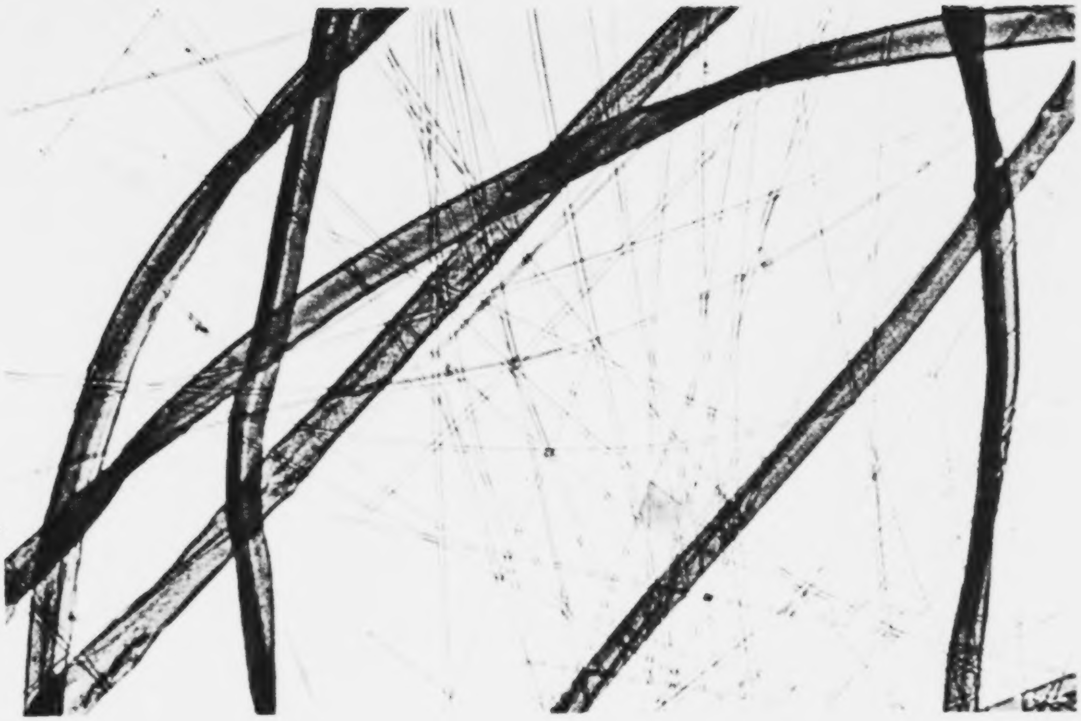


Figure No. 14 Photography by polarised light microscope of clean TEFAIRE^R needle felt 40 times enlarged.

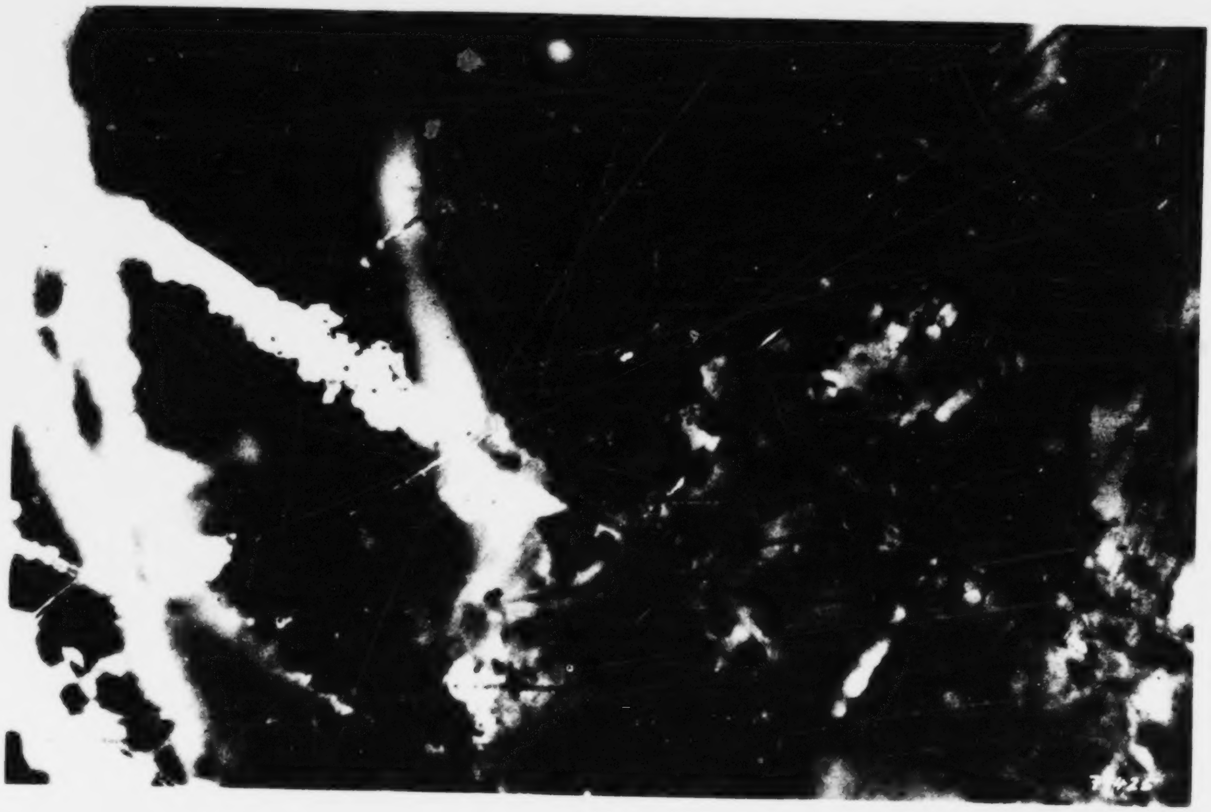
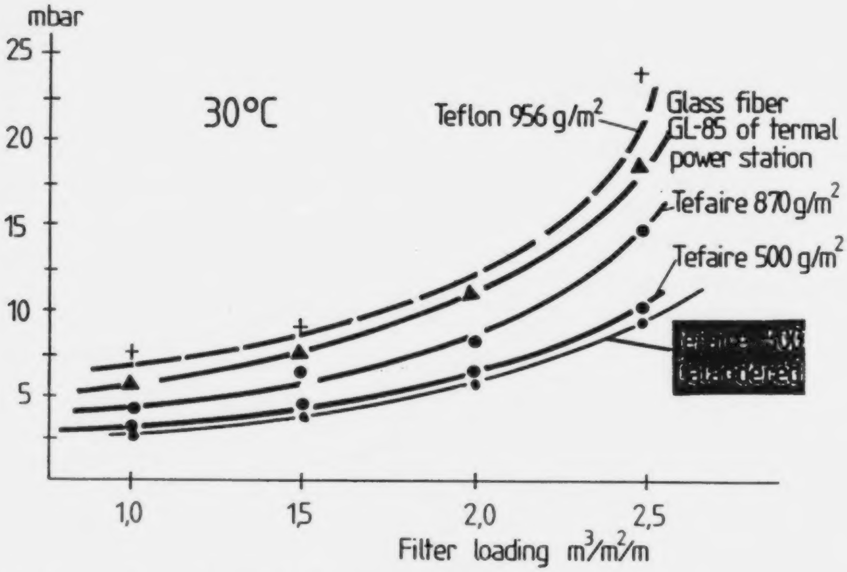


Figure No. 15 Photography by polarised light microscope of dusty TEFAIRE^{R} needle felt 40 times enlarged.

Material:	Fly ash of Braunschweig-Mitte power station Particle size 0,8-250 microns bulk density 0,96g/ml	
Parameters:	Dust loading	$50g/Nm^3$
	Bag cleanoff Cycle	1min 35secs.
	Measuring time	3hrs for each point
	Temperature	30°C



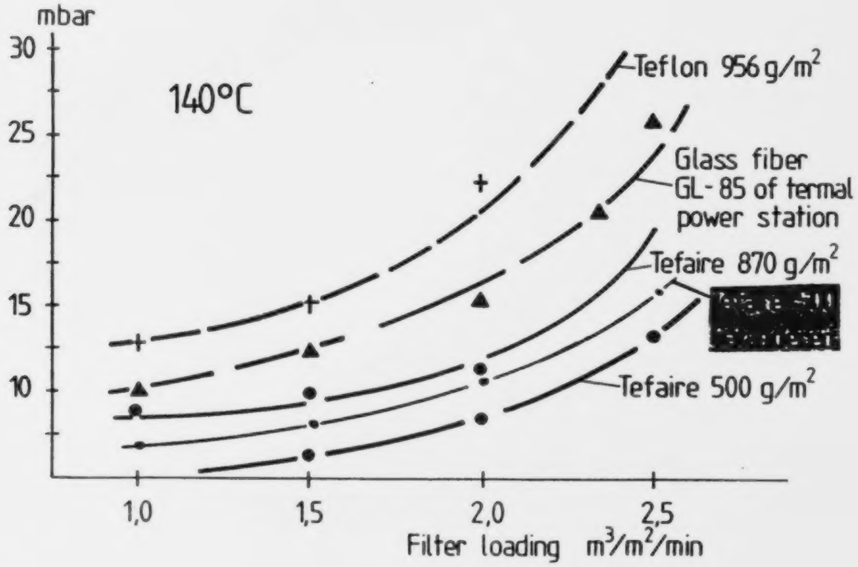
Test system: PGFB-9/24, boiler house, Uzwil plant

BUHLER

Table No. 2 Pressure differences as a function of gas-to-cloth ratios for different needle felt medias at 30°C (67° F)

Material: Fly ash of Braunschweig-Mitte
power station
Particle size 0,8-250 microns
bulk density 0,96g/ml

Parameters:	Dust loading	50g/Nm^3
	Bag cleanoff Cycle	1min 35secs.
	Measuring time	3hrs for each point
	Temperature	140°C

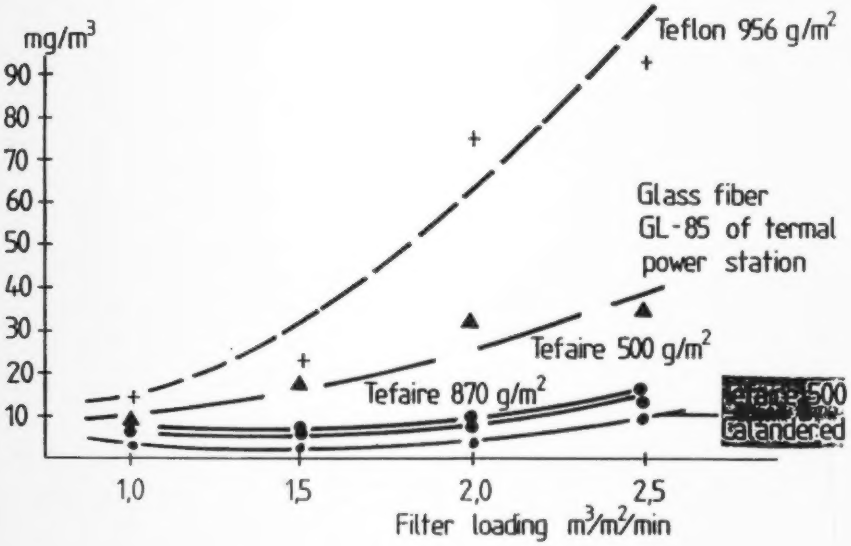


Test system: PGFB-9/24, boiler house, Uzwil plant

BÜHLER

Table No. 3 Pressure differences as a function of gas-to-cloth ratios for different needle felt filtermedias at 140°C (310°F)

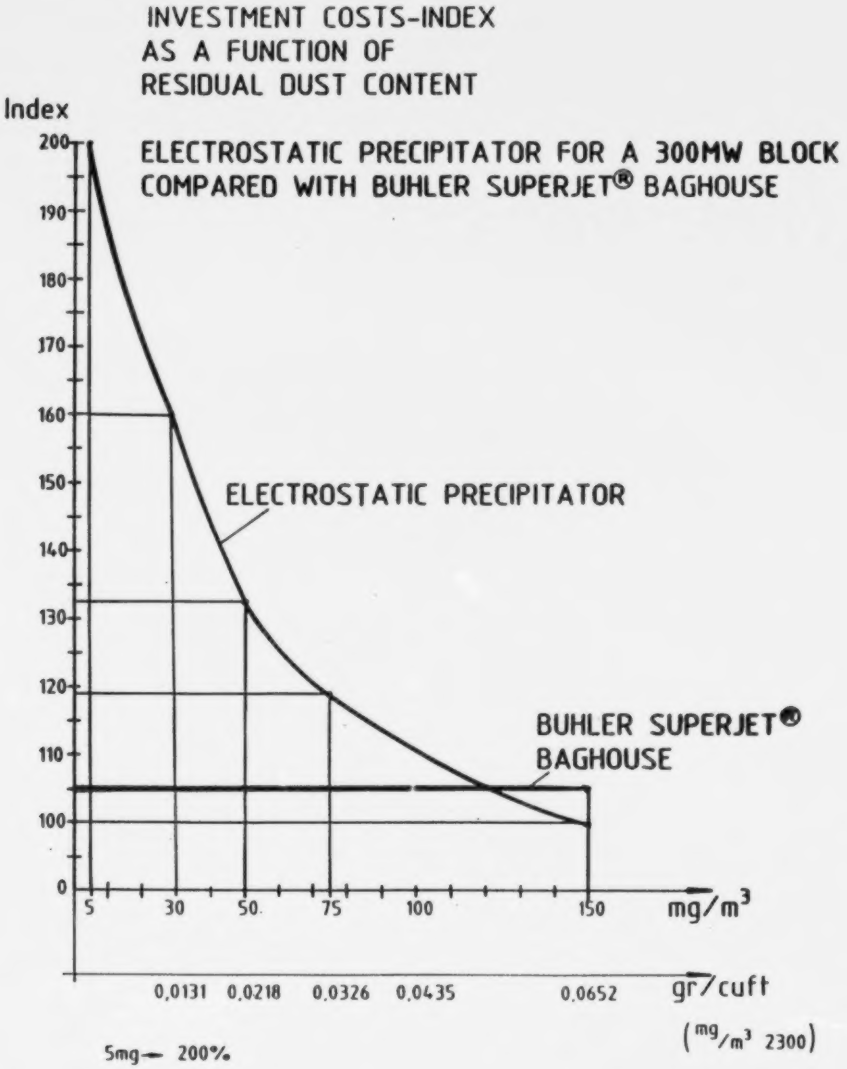
Material:	Fly ash of Braunschweig-Mitte power station Particle size 0,8-250 microns bulk density 0,96g/ml	
Parameters:	Dust loading	50g/Nm^3
	Bag cleanoff Cycle	1min 35secs.
	Measuring time	3hrs for each point



Test system: PGFB-9/24, boiler house, Uzwil plant

BÜHLER

Table No. 4 Residual dust content as a function of gas-to-cloth ratios for different needle felt medias at 30°C (67° F)



Umwelttechnik Transparent 1986 VDK/VIK 85 DPS/T20

Table No. 5 Investment cost index as a function of residual dust content in the clean gas.

ADVANTAGES OF BAG FILTERS: REFUSE INCINERATION

ELECTROSTATIC PRECIPITATORS --->	BAGFILTERS
USED TO BE ONLY POSSIBILITY OF SEPARATING DUSTS AT HIGH TEMPERATURES	DEVELOPMENT FABRIC
	NEEDLE FELT 140°C POLYESTER 180°C NOMEX 240°C GLAS FIBER 260°C TEFLON
	(RAYTON, PBI) TEMPERATUR-RESISTANT RESISTANT TO CHEMICALS
SERVICE LIFE OF ELECTRODES: GOOD (NO EFFECT DURING STARTING)	SERVICE LIFE OF BAGS: 3-5 years (effect during starting)
COSTS INCREASE	COSTS DECREASE
EASY MAINTENANCE: DIRTY GAS SIDE	EASY MAINTENANCE: CLEAN GAS SIDE
POOR SEPARATING EFFICIENCY	MAXIMUM SEPARATING EFFICIENCY
POORER EFFICIENCY IN SEPARATING THE DESULFURIZED GASES	DUSTS ARE FAVORABLE AFTER ADSORPTION PROCESS
DUSTS ARE UNFAVORABLE AFTER ADSORPTION OF THE ACIDS AND HEAVY METALS	

Table No. 6 Arguments for fibre filtration Versus electrostatic precipitators.

Friedrich Löffler (editor)
Hans Dietrich
Werner Flatt

Dust Collection with Bag Filters and Envelope Filters

With 203 figures

Translated by Hermann Simon

Friedr. Vieweg & Sohn Braunschweig / Wiesbaden

Table No. 7

New edition:
DUST COLLECTION with BAG FILTERS and ENVELOPE FILTERS
by Löffler, Dietrich, Flatt,
Friedr. Vieweg und Sohn Braunschweig/Wiesbaden

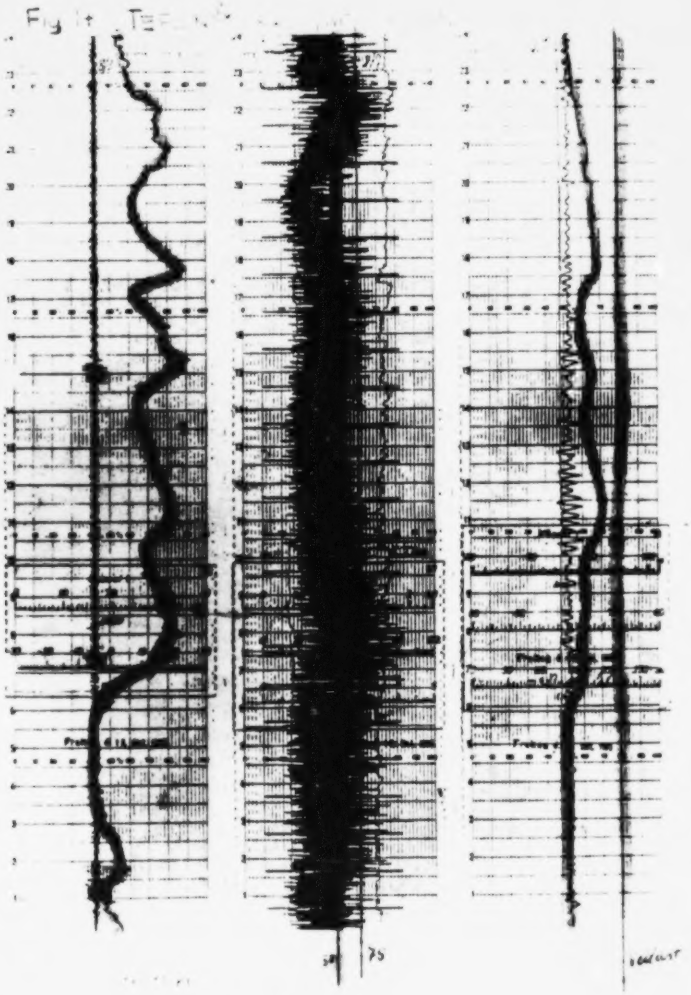


Table No. 8 Strip-chart record control room:
Differential pressure, residual dust content and boiler load when operating the plant with TEFLON^R filter bags.

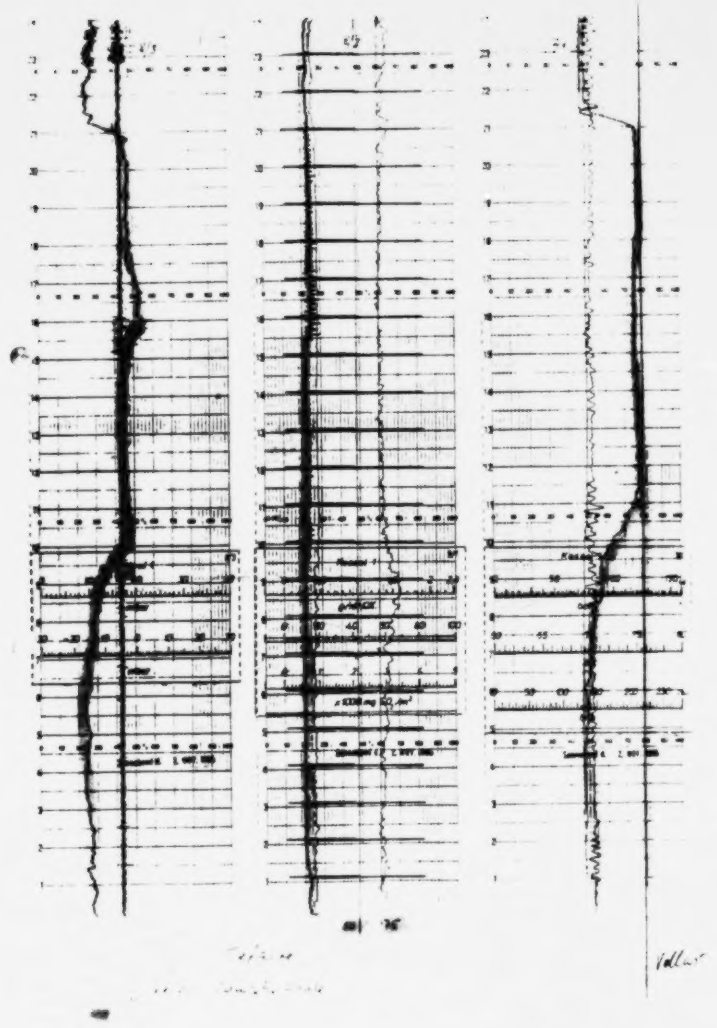


Table No. 9 Strip-chart record control room: Differential pressure, residual dust content and boiler load when operating the plant with TEFAIRE^R filter bags.

OPTIMIZING AND CONTROLLING BAGHOUSE OPERATIONS AT BG&E C.P. CRANE STATION

by:

John E. Sturtevant
Baltimore Gas and Electric Co.
Baltimore, Maryland

and

Gary P. Greiner
ETS, Inc.
Roanoke, Virginia

ABSTRACT

This is a summary paper of BG&E baghouse operation and maintenance activities from start-up to present, showing the application of modern troubleshooting and monitoring techniques in optimizing and controlling operation. Included are the following:

- Description of station and baghouses.
- Summary review of early operating history through end of vendor pilot plant work.
- Discussion of a comprehensive bag procurement program.
- Introduction of Individual Bag Flow Monitor (IBFM™) as a means of baghouse troubleshooting and evaluation of alternative fabrics.
- Use and experience of a Baghouse Performance Monitor (BPM™).

The work described in this paper was not funded by the U. S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

INTRODUCTION

This is a summary paper of the Baltimore Gas and Electric (BG&E) operating experience on its two baghouses at the C. P. Crane Station. It will describe the installation, discuss its unique characteristics and operational problems, and describe the procedures adopted to monitor and optimize operations.

SITE DESCRIPTION

The two steam generators at C. P. Crane Station are Babcock and Wilcox cyclone fired boilers which were placed in service in the early 1960's. No. 1 Boiler is a once through sub-critical design, whereas No. 2 Boiler is a drum type radiant design. Both boilers deliver 1,360,000 pounds of steam per hour at 2475 psi and 1050° F to turbine/generator units rated at 200 MW each. The cyclone boilers were converted from coal firing to burning No. 6 oil in the early 1970's due to the inexpensive oil prices and the impact of new strict air quality requirements. Oil firing continued at Crane until 1983 when the plant was converted back to coal. The conversion work included the addition of I.D. fans, changing the boilers from forced to balanced draft, and the construction of baghouses to replace the original inefficient electrostatic precipitators. The General Electric Environmental Services, Inc. (GEESI) baghouses have 10 compartments, each containing 522 bags which are 1 ft. in diameter and 35 feet in length. The bags are cleaned by reverse air. The baghouses were designed to filter the 835,000 ACFM full load gas volume with a flange-to-flange pressure drop of 6 inches W.C. with two compartments out of service. The baghouse inlet temperature is maintained at a minimum of 300° F by the use of supplementary steam coils located ahead of the Ljungstrom air heaters. The key design parameters are shown in Table 1.

EARLY HISTORY

The Unit #2 baghouse was placed in service on 5/13/83 followed by Unit #1 baghouse on 7/23/83. A high and unstable pressure drop was observed on the Unit #2 baghouse within 48 hours from start-up. Immediate steps to control the pressure drop were taken and included a continuous cleaning cycle, operation of two reverse air fans, and the addition of two sonic horns per compartment. These steps produced pressure drop stability but at a level that prohibited full load operation.

A corrective action program to reduce the pressure drop to guarantee levels was undertaken by the baghouse vendor. This was a multipronged investigation utilizing lab simulation, a field pilot baghouse, and full scale evaluations. Details of this 2 year effort can be found in the Shilling, et al paper presented at the 3rd Conference on Fabric Filter Technology for Coal Fired Power Plants, Scottsdale, Arizona in November, 1986. The key outputs of this program were:

1) Utilization of four sonic horns per compartment.
2) Enlargement of outlet poppet damper diameters on the first four compartments.
3) Enlargement of the outlet duct.
4) Modification of the reverse air duct.

Table 2 compares the Unit #2 pressure drop performance of November, 1984, following the addition of four sonic horns per compartment, with the performance of April, 1985, following all mechanical modifications. The overall pressure drop reduction resulting from the mechanical modifications was approximately 1.0 inches W.C.

Unit #1 did not see the severe early ΔP problems as Unit #2, but also failed to meet the guaranteed ΔP. To minimize pressure drop in this baghouse, the same modifications were implemented.

The conclusion of this effort was that the baghouse systems, as installed, are the best that can be achieved given the special inlet conditions from the Crane Station cyclone boilers. Specifically, it concluded that the high ΔP is due to the very fine ash particles, which develop a residual dust cake that has high flow resistance and is difficult to remove. This in turn dictates a short filtration cycle, which coupled with the lower inlet mass loading generated by the cyclone type boiler, yields an order of magnitude less dust per filtration cycle (see Table 3). This reduces the effectiveness of bag cleaning as there is not enough released dust within the bags to generate the plug action required to both scour the bag surface and induce the inward motion and cleaning air flow as shown in Figure 1.

Based on GEESI's two year investigation and modification program and the April, 1985 test results, the baghouses were accepted by BG&E.

The baghouses ran in a stable manner through the summer of 1985. During this period, Unit #1 was restricted to 160 MW due to excessive (50%) air preheater leakage which increased the gas flow to the baghouse. The baghouse inlet temperature was maintained above 300° F. When the ambient temperature dropped, in late fall, the baghouse inlet gas temperature fell to as low as 270° F. Pressure drop and bag weights began to rise. An outage was taken to repair the air preheater seals in order to reduce leakage. During this outage, the bags were reverse flushed with ambient air and each bag was manually shaken in an attempt to remove the heavy dust cake.

Following start-up, the air preheater leakage was 25% and the baghouse full load pressure drop ran 4-5 inches. Within one week the pressure drop increased to over 9 inches. Load reduction was performed to maintain a maximum 9 inch ΔP. Within two weeks, the load was restricted to 125 MW.

EQUIPMENT OPTIMIZATION

In December, 1985, ETS, Inc., a pollution control consulting firm from Roanoke, VA, was contracted to address the pressure drop problems in Unit 1. During the initial site visit, operational data strongly suggested that insufficient cleaning energy was reaching the bags. It was discovered that one of the module reverse air poppet dampers had failed open, thus stealing cleaning air from the other nine compartments and reducing cleaning effectiveness. Correction of this problem immediately reduced the pressure drop 1 1/2 inches W.C. at 125 MW and allowed the unit to be operated at 180 MW, but still at abnormally high ΔP levels.

Fabric lab analyses showed a significant loss of strength had occurred since the low temperature excursions. The conclusion was reached that the present bag set had been permanently blinded and degraded by the low temperature operation during the period of high air preheater leakage. At this point, the plant decided to attack the problem with a comprehensive program. The goals were to place the baghouse in a condition allowing full load operation, and the development of instrumentation which would continuously monitor the baghouse performance and provide an early warning of filtering system problems. In order to achieve this, it was decided to:

1) Replace the bag set.
2) Develop an alternative fabric testing program.
3) Install a baghouse monitoring system.

BAG SET REPLACEMENT

The first step was to obtain a high quality replacement bag set. In order to properly procure filter bags in an orderly and timely manner, without last minute compromises of quality and delays in delivery, a program including the following ingredients was required:

1) A predefined fabric specification, which would be agreed upon in advance of any production or testing.

2) Finished rolls of fabric to be released for bag fabrication based on a predefined Q.A. test program performed by an independent testing lab.

3) A bag fabricator contracted to fabricate a bag set to an agreed upon bag specification, utilizing the released fabric, with delivery within 2-3 weeks following authorization to proceed.

4) A quality audit program during bag set fabrication by a BG&E representative or independent firm.

5) Development of a bag installation procedure insuring a high quality installation technique while minimizing

unit load reductions.

6) Contracting an experienced bag installer to perform the work as specified.

In today's world, this requires substantial effort and time as shown in Figure 2. A six to seven month lead time can be expected.

Selection of the bag fabric was difficult due to lack of experience with alternate fabrics in the Crane baghouses. Experience to date had clearly shown that the Crane baghouses were unique from all other baghouse sites making the selection of a new type fabric a risky proposition. It was decided to use the same fabric as the initial bag set for the replacement bags since two and one half years of acceptable performance had been obtained.

A complete specification was developed along with a quality audit plan to obtain a known and uniform bag set. The bag construction was defined by a detailed drawing (Figure 3) and a specification document that defined the requirements for fabric, sewing thread, bag hardware, bag construction, marking, packing, and shipping. The specification also defined the procedures and acceptance criteria used to sample and release fabric and filter bags. The ETS fabric testing lab was used to perform a quality audit. This consisted of testing and releasing fabric headers before fabrication and periodic quality checks during fabrication.

This program produced the desired high quality bag set and resulted in only six bags being rejected during installation. Based upon the experience from rebagging Unit #1, a list of recommendations was developed for future BG&E rebagging projects and is shown in Table 4.

ALTERNATIVE BAG EVALUATION

A bag testing program was developed for the purpose of assessing alternative bag types for future use in the Crane baghouses. Seven bag types were selected as being viable candidates. Twenty-four bags of each type were purchased and installed in the baghouse. Twelve bags of each type were installed in compartment #120 as the main test location and the other twelve bags of each type in compartment #170 as the back-up location. The test bags were permanently tagged and positioned in specific locations for positive future identification as described in Figure 4.

Twenty-four Individual Bag Flow Monitors (IBFM™'s)* and associated instrumentation were purchased from ETS, Inc. and installed in the test compartments. The IBFM™'s were installed on the bag mounting thimbles, Figure 5, to measure the actual gas flow through the bag. This allows monitoring of the individual bag gas-to-cloth ratios and drag levels providing a side by side comparison of the flow characteristics of the various test bags. Flow performance data on the test bags are periodically collected and trended. In addition, test bags are removed on a scheduled basis and sent to a testing lab for strength analysis and visual inspection. In this manner, experience is being gained with the alternative bag types under the actual operating conditions in the Crane baghouse. This knowledge will enable BG&E to make future selections for replacement bags which provide the best combination of bag life and pressure drop performance.

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*U. S. Patent No. 4,719,791

The alternative fabrics have been monitored through 19 months of operation as of February, 1988. Periodically, the IBFM™ system has been used to define the G/C and drag of each bag type through several filtration cycles so that comparison of average G/C, residual drag, average drag and K_2 can be made. Figure 6 shows the relative fabric drag levels recorded at the midpoint of a filtration cycle. Because of the short filtration cycles and continuous cleaning mode, this midpoint drag can be utilized to project what the total baghouse pressure drop would be if all bags were of that type. Table 5 shows the relative projected pressure drop and impact on annual I.D. fan costs. As can be seen in both Figure 6 and Table 5, two bag types stand apart from the others in flow performance and would add a significant operating cost. Since the baghouse cannot operate at full load above 9" ΔP, types 3 and 5 are not suitable candidates for future bag sets.

Figure 7 shows the comparative trend results of one of the typical strength parameters. Similar graphs have been prepared for all lab tests including: Permeability, Mullen Burst, Tensile, MIT Flex, and Weight.

So far, all the fabric types appear to have stabilized at strength levels dictated by their construction. Only continued monitoring through the practical life of the bags will determine which fabric types is best for Crane. Types 1, 2 and 4 were expected to exhibit similar performance and to date they have. Type 6 is an experimental type fabric. Type 3, reported to have provided lower pressure drop at a Texas Utility and Type 5, used at several sites to extend bag life, impose a severe pressure drop penalty at Crane. Type 7, included with the promise of lower pressure drop, but at a significantly higher bag cost, has not shown any superior flow characteristics. Experience to date further emphasizes the danger of translating information from one utility site to another.

BAGHOUSE PERFORMANCE MONITORING

The installation of IBFM™'s also provides a means of monitoring the general baghouse condition. A continuous Baghouse Performance Monitoring System (BPM™) was developed by ETS, Inc. and installed on the Crane #1 baghouse. This system is designed to alert the plant of subtle changes in the baghouse performance. This early warning will allow corrective actions to be taken before the problem becomes major, thus averting load reductions and/or significant bag damage.

The BPM™ system samples, calculates and prints out a continuous recording of G/C and system ΔP levels adjusted to full load conditions as shown in Figure 8. The current method of trending requires manual strip chart analysis and data reduction. Future BPM™ equipment is being designed to provide digital datalogging of the monitored parameters and remote modem access for analysis and data reduction. Trending information is plotted as shown on Figure 9. If the system is operating in a normal and stable manner, the traces will be within a flat band. If either or both traces trend outside the normal band, this is a signal to check the system operation. Too high a G/C level will indicate increased air preheater leakage or abnormally high boiler O_2 operation. A higher pressure drop without corresponding higher G/C will indicate loss of cleaning energy or change in inlet conditions. Long term gradual drifts will show the magnitude of the rate of the natural bag set aging process and assist in projecting bag life. Such a long term trend report is shown in Figure 10.

When baghouse problems are indicated by a rise in ΔP without a corresponding rise of gas flow, the BPM™ system can be used as a diagnostic tool to help troubleshoot the cause. Individual filtration and cleaning cycles can be examined to detect the proper operation of dampers and the required level of cleaning air. This information can be captured on the BPM™ strip chart and is simulated on Figure 11. This technique has been used at Crane to confirm the cause of problems and assist in isolating the source.

CLOSING

The early baghouse experience at C. P. Crane Station was one of uncertainty and frustration for both the vendor and the plant. It was a learning experience for everyone involved. Now that a basic understanding has been acquired of the difficulties encountered in filtering ash from a cyclone boiler, the plant is moving toward optimizing the operation of the existing equipment. It has been shown that the baghouses can perform their intended job. However, the systems must be maintained in nearly perfect working order. Close monitoring of the baghouse performance and the early identification of problems are the essential elements in reducing the frequency of baghouse related load reductions.

FIGURES

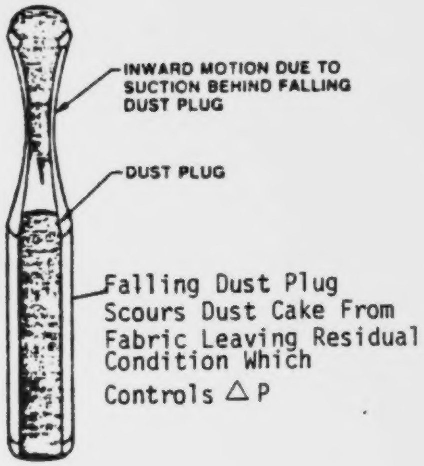
INWARD MOTION DUE TO SUCTION BEHIND FALLING DUST PLUG
DUST PLUG
Falling Dust Plug Scours Dust Cake From Fabric Leaving Residual Condition Which Controls △P

Figure 1. Plug Mode of Cleaning

		MONTHS										
TASK	1	2	3	4	5	6	7	8	9	10	11	12
1) Produce Fabric & Bag Specifications	X											
2) -Solicit Bids	XXXXX											
-Select Bag Vendor		X										
3) -Obtain Fabric			XXXXXXX									
-Q.A. Testing					XXX							
-Release to Storage					XXXX							
4) -Release Bag Fabrication						X						
-Delivery to Plant						XXX						
5) -Procure Bag Installers				XXXXXXXX								
-Install Bag Set						XXXX						

Figure 2. Bag Procurement Task Schedule

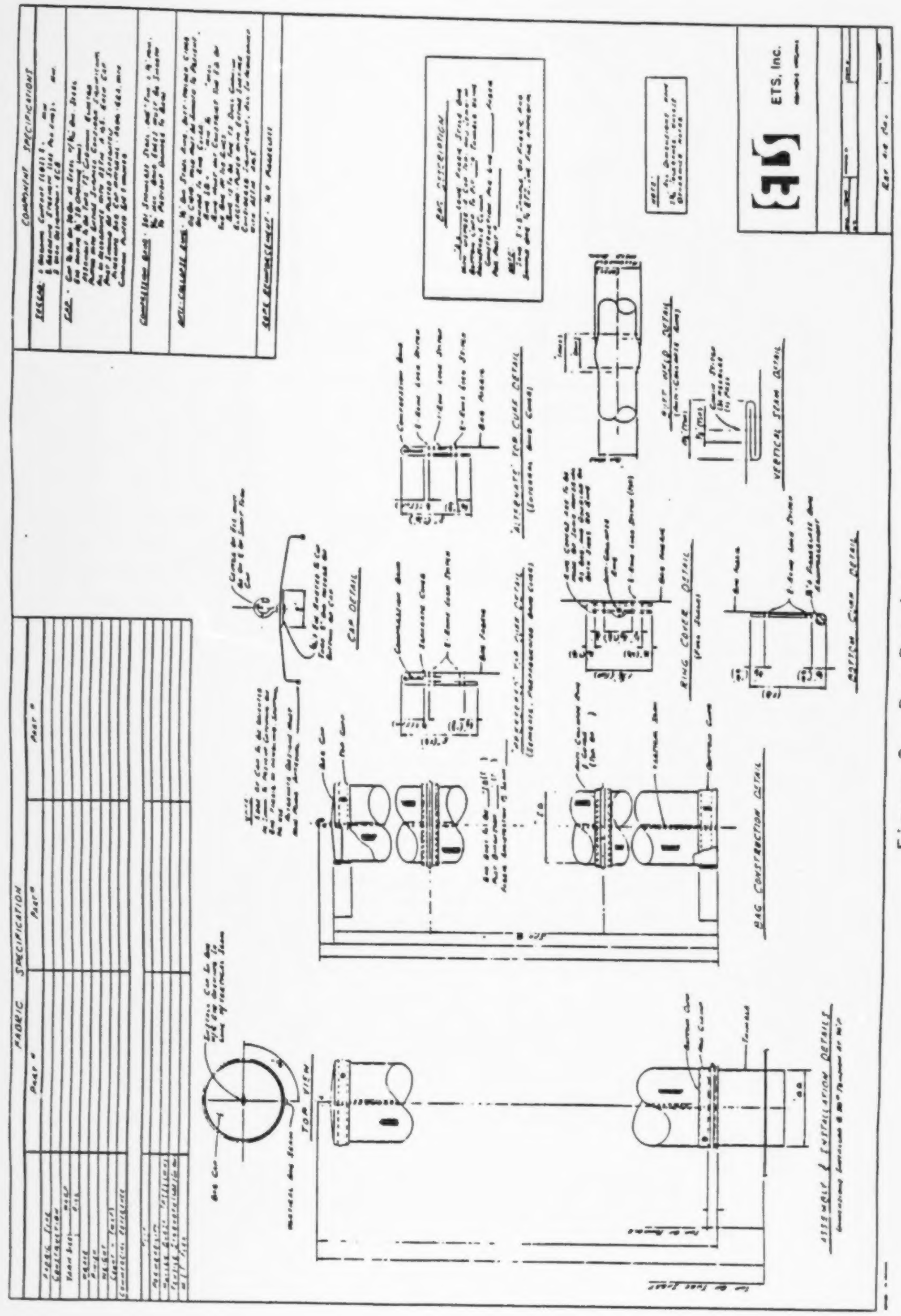

Figure 3. Bag Drawing

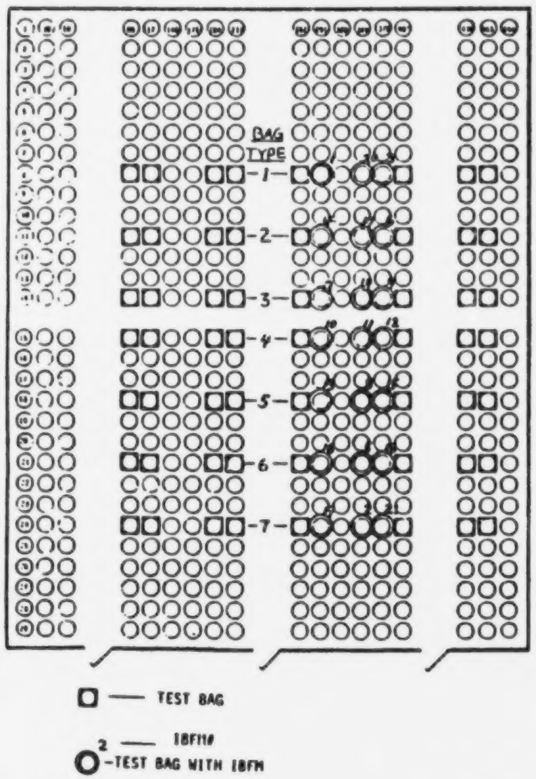

Figure 4. Test Compartment Layout

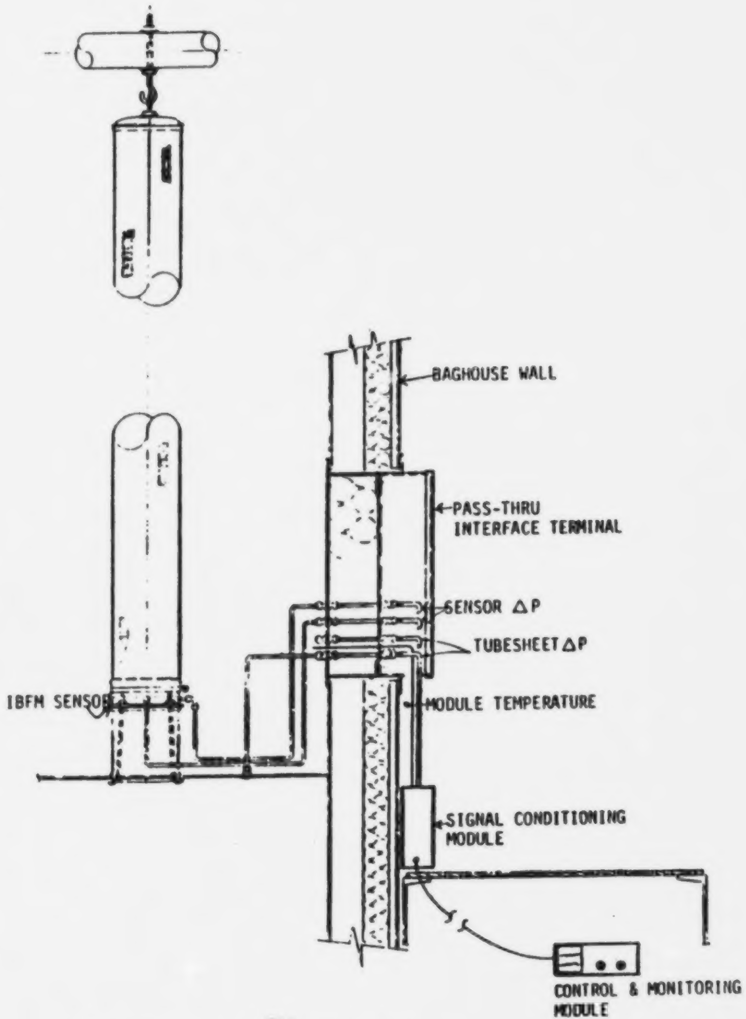

Figure 5. IBFM^{TM} Installation

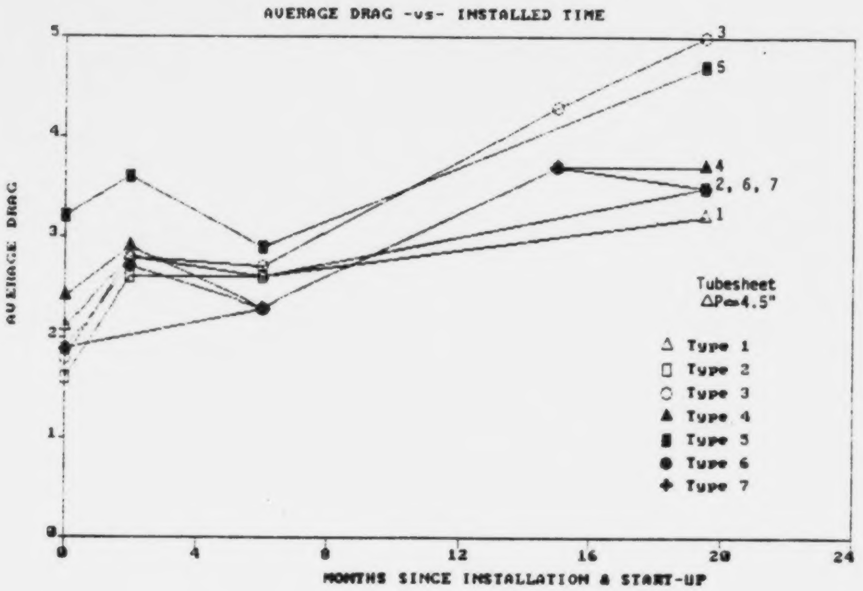

Figure 6. Unit #1 Bag Matrix Flow Performance vs. Installed Time

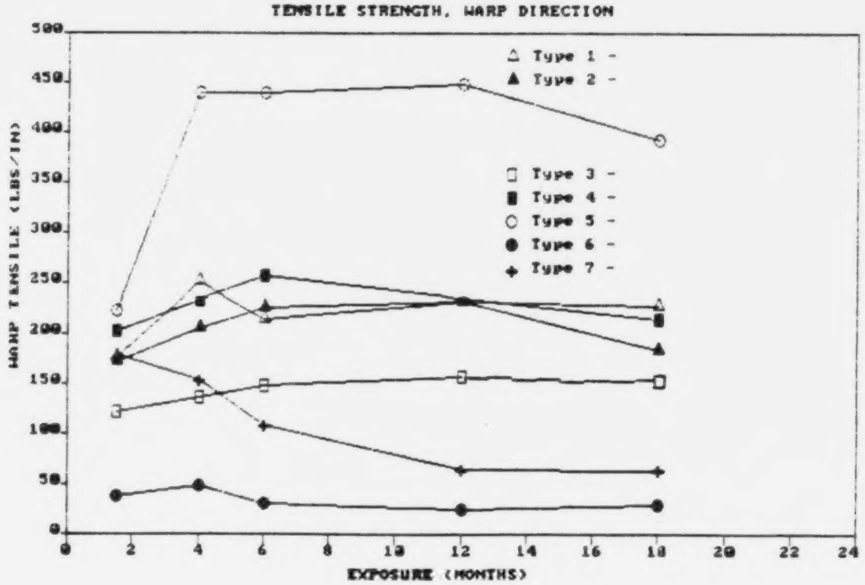

Figure 7. Unit #1 Bag Matrix Warp Tensile Strength vs. Installed Time

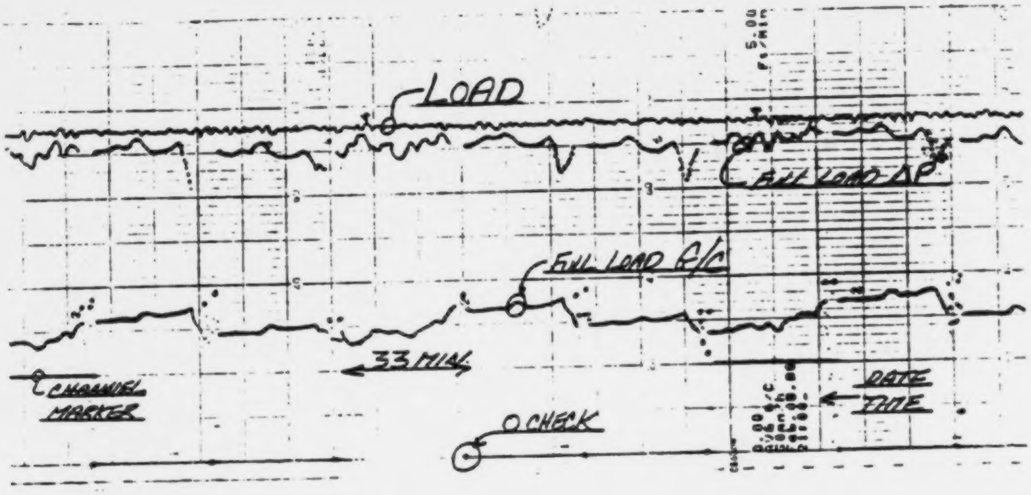

Figure 8. Typical BPM Strip Chart Output
Standard Monitoring Mode

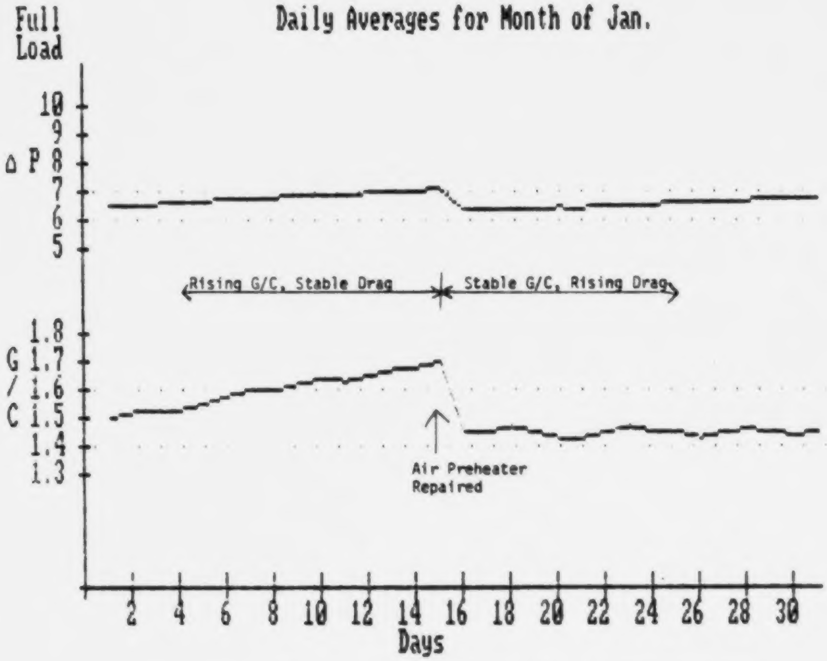

Figure 9. Typical Monthly Report Full Load G/C and
Δ P vs. Time

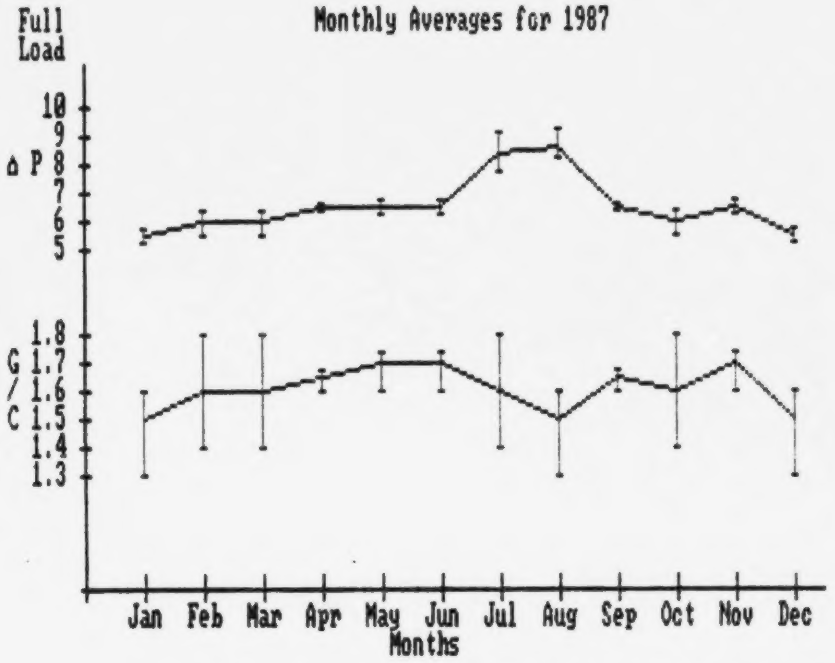

Figure 10. Typical Monthly Trend Report Full Load G/C, and △P Daily Averages

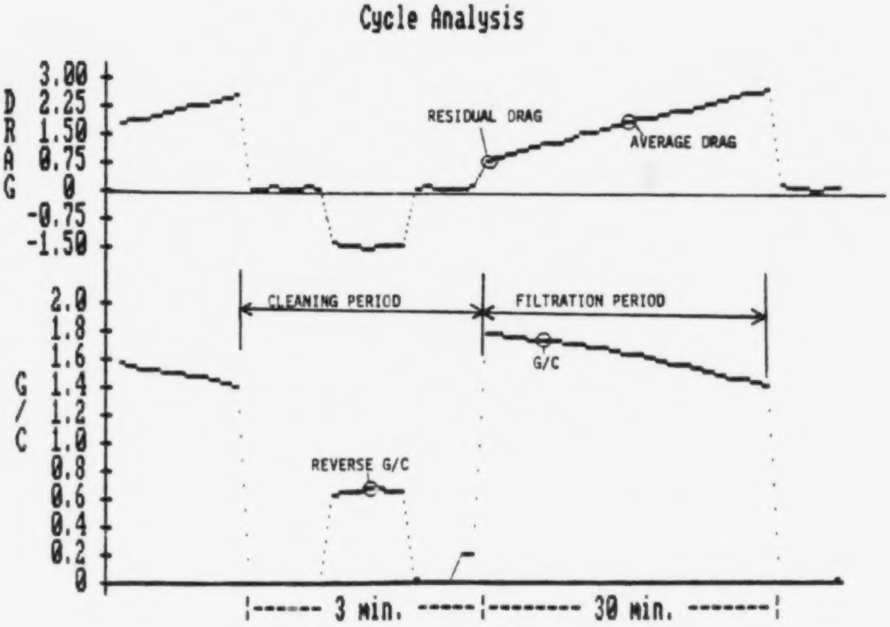

Figure 11. Troubleshooting with the BPM Cycle Analysis

TABLES

TABLE 1. KEY DESIGN PARAMETERS OF C. P. CRANE UNITS 1 OR 2 BAGHOUSE

Item	Design
Baghouse Model	GE #10-522-12-35
No. Modules	10
No. Bags per Module	522 - 12" Ø x 35' Lg
Total Cloth Area (Ft^2)	574,200
Full Load (MW)	200
Full Load Gas Vol. (ACFM)	835,000
Baghouse Inlet Temp. (°F)	300
Rev. Air Volume (ACFM)	92,044
Inlet Loading (gr/ACF)	1.27

Operating Performance	G/C	ΔP
8 Modules w/o R.A.	1.82	6.0 (Guaranteed)
9 Modules w/o R.A.	1.62	5.75
9 Modules w/ R.A.	1.79	—
10 Modules w/o R.A.	1.45	5.50
Reverse Gas Flow	1.60	

Bag Type - 9.50 oz/sq. yd - 3 x 1 Twill Woven Glass with Acid Resistant Finish

I.D. Fan Power Operating Cost - $50,000/inch/yr

Bag Set Cost (Installed) - $500,000/set

TABLE 2. BEFORE AND AFTER MECHANICAL MODIFICATIONS - UNIT #2

	Flange-to-Flange ΔP @ 835,000 acfm Inlet	
	11/84	4/85
10 Compartments w/o RA	6.28	5.54
9 Compartments w/o RA	7.00	6.16
9 Compartments @ RA = 135,000 ACFM	8.63	7.58
9 Compartments @ RA = 92,500 ACFM	—	7.12

TABLE 3. IMPACT OF UNIQUE CRANE PROPERTIES ON MASS DUST COLLECTED DURING ONE FILTRATION PERIOD

SITE	TYP G/C	MASS LOADING GR/ACF	MASS MEDIAN PART. SIZE MICRON	FILT. CYCLE TIME MIN.	K_2 (IN. H_2O/FPM/(LB/FT^2))	MASS DUST* PER FILT. CYCLE LBS/FT^2
BG&E Design	1.46	1.27	—	—	31.16	—
Actual Unit 1	1.6	0.8-1.0	9-10	33	24-60	6.8×10^{-3}
Actual Unit 2	1.6	0.32-0.5	6.5-8	33	—	3.0×10^{-3}
Typical PC Site	2.0	2-5	15-20	90-240	7-10	100×10^{-3}

*lbs/ft^2 = G/C x Filtration Cycle Minutes x $\frac{GR/ACF}{7000}$

TABLE 4. BAG PROCUREMENT AND BAG INSTALLATION RECOMMENDATIONS FOR BG&E CRANE STATION BAGHOUSES

PROCUREMENT

1) Select the type of bag to be installed well in advance.
2) Solicit bag bids and place order three months prior to he anticipated required delivery date.
3) Contract an independent bag testing lab to perform fabric testing to assure adherence to the specification.
4) Visit the vendor's plant to witness bag fabrication and check for compliance with their quality assurance program.

INSTALLATION

1) The rebagging should be done while the unit is in service. This minimizes the risk of damage to the new bags when they are initially placed in service.
2) The summer months should be avoided if rebagging while in service.
3) The installation specification should specifically address the possibility of high compartment temperatures (up to 150°F) and heavy bag weights (approximately 100 pounds).
4) The reverse gas duct pressure should be closely monitored and maintained at a negative pressure.
5) A continuous oxygen monitor should be used to detect the presence of flue gas.
6) Full face respirators should be worn during the bag stripping operation.

TABLE 5. COMPARISON OF BAG TYPE IMPACT ON BAGHOUSE FL-FL PRESSURE DROP AND ANNUAL I.D. FAN COST

BAG TYPE	AVG. FL-FL ΔP (INCHES W.C.)	ANNUAL I.D. FAN COST FROM CONTROL BAG ($000'S)
1	7.1	- 40
2	7.6	- 15
3	10.2	+115
4	7.9	0
5	9.5	+ 80
6	7.6	- 15
7	7.6	- 15

- Avg. FL-FL ΔP = 19 month average drag x 1.6 (G/C) + 2"
o I.D. Fan Cost Based upon $50,000/yr/inch ΔP

PREDICTING RELATIVE DRAG FOR BAGHOUSES FROM COAL CHEMISTRY

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ABSTRACT

Filtering drags and average dustcake areal densities measured at nine baghouses collecting pulverized-coal fly ash and employing reverse-gas cleaning and filtering fabrics with 75% exposed surface texturization have been compared with dustcake ash and coal ash chemistry. Strong correlations have been identified relating chemical compounds in the dustcake ashes to filtering drag and average dustcake areal density. Correlations were also identified between the measured concentrations of several chemical compounds in the coal ashes and corresponding concentrations in the dustcake ashes. Finally, correlations relating coal ash chemistry to baghouse filtering drag and dustcake areal density were identified. These relationships allow filtering fabric and cleaning method to be tailored to the chemistry of the source coal.

INTRODUCTION

Baghouse design and operation, specifically selection of cleaning method and fabric, can be tailored to match the characteristics of the ash being collected. A wide range of baghouse performance has been encountered at baghouses collecting fly ash generated from pulverized-coal (PC) applications (1,2,3). For the purposes of this paper, we define baghouse performance in terms of dustcake areal density and normalized drag (Normalized drag is calculated by dividing the measured average tubesheet drag by the dustcake areal density.) This paper does not address collection efficiency or bag life. The ultimate objective of this paper is to present empirical relationships for the prediction of dustcake weight and normalized drag based on the chemical characteristics of the coal to be burned.

To determine if this objective was feasible, we initially looked at correlations between baghouse performance data and dustcake ash chemistry. These comparisons revealed several chemical components of the dustcake ashes that correlated well with baghouse performance. These chemical components of the dustcake ash were compared with chemical characterizations of source coal samples. Some compounds in the dustcake ash were found to relate strongly to corresponding compounds in the source coal. Direct comparisons were then made between the baghouse performance data and the coal chemistry. This third comparison step has yielded the desired empirical relationships for the prediction of dustcake areal density and normalized drag for the applications described in this paper.

This effort is analogous to the development of the empirical relationships for the prediction of fly ash resistivity and estimation of electrostatic precipitator performance. Relating baghouse performance to coal chemistry provides the means to tailor baghouse design (including selection of cleaning method and fabric) for specific coals.

ASSEMBLY OF DATA

Empirical relationships were developed through the assembly of data describing baghouse performance, the chemistry of the fly ash composing the filtering dustcake (dustcake ash chemistry), and the chemistry of the source coal (coal chemistry). Empirical relationships are limited in their application to the range of data from which the relationships were derived. The data in our database represent baghouses collecting fly ash at PC plants burning coal mined in the United States.

BAGHOUSE PERFORMANCE DATA

We have assembled data describing representative performance for nine full-scale and pilot-scale baghouses. These data are summarized in table 1, and are representative of baghouse performance with mature dustcakes, reverse-gas cleaning (without sonic assist), and fabric with 75% exposed surface texturization. (Some of the plants represented have altered their baghouse operation since these data were acquired.)

The size distribution of the particles composing the filtering dustcake can have a strong effect on drag (4,5). The presence of a multiclone or other mechanical collector upstream of a baghouse can significantly reduce the median size of the ash particles reaching the dustcake. The normalized drag of this dustcake can be greatly increased. A significant change in median particle size may also alter ash cohesivity. None of the data presented in this paper represent operation with mechanical collectors, though some studies have been performed to quantify the effects of this type of operation.

Filtering air-to-cloth values were calculated from measurements obtained with orifice plates, venturis, or pitot traverses of inlet ducts. Dustcake areal density values were obtained by measuring bag weights in situ (immediately following cleaning cycles). Average dustcake areal density in the tested compartment is shown, unless the actual areal density in the tested bag was available.

DUSTCAKE ASH CHEMISTRY

Chemical characterizations of dustcake ashes from the nine operating baghouses are also presented in table 1. When multiple samples were available from a given baghouse, the average chemical profile of dustcake ash obtained from that baghouse is presented. (Correlations presented later in this paper are based on these averaged values.) Consistent methods of sample acquisition and data measurement were of prime importance in the assembly of the dustcake ash data. The data characterizing the chemistry of the dustcake ashes were measured in the laboratories of Southern Research Institute.

COAL CHEMISTRY

Laboratory analyses were performed on representative coal samples obtained from the plants represented in table 1. These coal samples were subjected to proximate and ultimate analyses, as well as analysis of the mineral constituents of ash produced by ignition of the coal samples in a 750°C oven (coal ash). Table 1 includes mineral analyses of the coal ash samples and the measured ash and sulfur contents of the source coal samples. When multiple coal samples were available from a given plant, the averaged data measured for coal samples from that plant are presented. (As with the dustcake ash chemistry data, correlations presented later in this paper are based on these averaged coal analyses values.)

CORRELATING DATA

The correlations of primary interest for predicting normalized drags and dustcake areal densities are those which will allow these predictions to be based on coal characteristics. However, the dustcake ash corresponds more directly than the coal ash to the baghouse performance data. Therefore, the first task was to determine if any significant correlations exist between the dustcake ash chemical characteristics and the baghouse performance data.

COMPARING BAGHOUSE PERFORMANCE DATA WITH DUSTCAKE ASH CHEMISTRY

The chemical characteristics most strongly correlated with baghouse performance data are plotted versus measured baghouse performance data in figures 1 through 6. These relationships are strong enough to allow reasonably reliable estimations of normalized drag and dustcake areal density to be based on chemical characterizations of dustcake ashes.

Effects of Alkalis on Cohesivity

Dustcake areal densities are higher at baghouses collecting ashes from high-sulfur coals (1). The data presented in figures 1, 2 and 3 relate calcium, magnesium and alkalinity (presented as the equilibrium pH) to dustcake areal density. These trends indicate that the cohesivity of fly ash (which directly affects dustcake areal density) is influenced by the presence of alkali compounds in the fly ash. A possible explanation of this effect is that at least some of the alkali compounds in the fly ash regulate the cohesivity of the dustcake by their reaction with the sulfur oxides in the flue gas.

The sulfur in the coal, volatilized during the combustion process, is available to react with alkali compounds. Calcium and the other alkalis in the fly ash can react with the gas-phase sulfur oxides, binding them up as non-cohesive sulfates. Unreacted sulfur oxides, in substantial concentrations in the flue gas from the combustion of high-sulfur, low-alkali coals, increase ash cohesivity (and subsequently dustcake areal density). Low-sulfur coals may contain sufficient alkali content to react with the sulfur oxides in the flue gas to keep ash cohesivity low. This reaction can occur in the furnace and in the gas passages, and on the dustcake.

Relationships supporting this mechanism are presented in figure 3, relating the equilibrium pH of the dustcake ashes to the measured dustcake areal densities, and in figure 7, which presents the positive correlation between the calcium and the sulfur in the dustcake ash. (This correlation is not evident in hopper ashes, indicating a significant part of the reaction occurs in the dustcake.) These relationships, along with others presented in this paper, support the conclusion that unreacted sulfur oxides have a strong influence on ash cohesivity, and that the levels of these unreacted sulfur oxides are primarily regulated by alkali content. Alkali content is therefore a primary indicator of ash cohesivity.

COMPARING DUSTCAKE ASH CHEMISTRY WITH COAL CHEMISTRY

Since baghouse performance can be estimated from empirical relationships based on dustcake ash chemistry, we sought correlations between the dustcake ash chemistry data and the mineral analyses of the coal ashes derived from the source coals. We found that components of the chemistry of the dustcake ashes can be predicted with reasonably high confidence from the mineral analyses of the coal ashes.

A strong degree of correlation was found between the concentrations of calcium and magnesium measured in the dustcake ashes with the concentrations of calcium and magnesium in the coal ash samples. The relationship observed for calcium (shown in figure 8) along with similar relationships for sodium, magnesium, and potassium indicate that the relative amounts of these compounds are not significantly altered by the differences in plant design and operation over the range of variations that were in the database. Other chemical components in the coal ash samples do not correlate as well with corresponding compounds in the dustcake ash samples.

COMPARING COAL ASH CHEMISTRY WITH BAGHOUSE PERFORMANCE DATA

Because the concentrations of calcium and magnesium in the dustcake ashes correlate well with observed baghouse performance data (dustcake areal densities and normalized drags), and also with measurable chemical characteristics of the source coal ashes, direct comparisons were evaluated between the chemical constituents of the coal ashes and baghouse performance data.

Comparisons of dustcake ash chemistry with baghouse performance data have revealed that dustcake areal density values are strongly correlated with the concentrations of calcium and magnesium in the coal ashes. Figure 9 presents the relationship between dustcake areal density values and the calcium in the coal ash samples. (The relationship between dustcake areal density and the magnesium in the coal ash samples is similar to that shown in figure 9.)

The concentrations of calcium and magnesium in the coal ash samples also correlate strongly with the normalized drag. The relationship between coal ash calcium concentration and normalized drag is presented in figure 10. (The relationship between coal ash magnesium and normalized drag is similar to that shown in figure 10.)

SUMMARY OF CORRELATIONS

Concentrations of calcium and magnesium in coal ash samples provide a means for predicting baghouse areal density and normalized drag, the components of filtering drag. The data in figure 9 indicate that baghouses collecting ashes derived from coals with relatively high concentrations of calcium (also true for magnesium) exhibit relatively low dustcake areal densities. That is, higher concentrations of calcium and magnesium in the coal (and subsequently in the dustcake ash) should always produce a less tenacious, easier-to-remove dustcake. Figure 10 illustrates the tendency of baghouses collecting ashes derived from coals with relatively high concentrations of calcium (also true with magnesium) to exhibit relatively high normalized drags.

Ash cohesivity (expressed in terms of dustcake areal density) is better indicated by calcium (or magnesium) content than by coal sulfur content. This is evident in the comparison of the data presented in figure 11 (relating coal sulfur content to measured dustcake areal density values) with the data in figure 9 for coal ash calcium (and similar data for magnesium). The extent of the effects of fabric choices and cleaning methods not represented in our database on dustcake areal density and normalized drag must be drawn from field experience. Other cleaning methods and filtering fabrics will alter dustcake areal density and possibly normalized drag (6); however, the relative rankings of these quantities, as indicated by coal ash chemistry, should remain the same.

USING THE EMPIRICAL RELATIONSHIPS

Areal density and normalized drag tend to offset each other's effects, as is evident in their inverse relationships to calcium and magnesium. Coals that form dustcakes with high areal densities usually have low normalized drags, and vice versa. This result is not unexpected, because of the effects of relative cohesivity on dustcake formation. High-cohesivity pulverized-coal ashes tend to form dustcakes that have a high characteristic porosity, and therefore generally exhibit low normalized drag (5). These same high-cohesivity ashes form relatively tenacious dustcakes with high areal densities. Conversely, low-cohesivity pulverized-coal ashes form relatively thin, relatively dense dustcakes with high characteristic normalized drag.

It has been shown that cohesivity is strongly related to the alkalinity of dustcake ash (expressed as the equilibrium pH). It is desirable to examine other quantities that can be used to express the alkalinity of coal ash. One possible expression is the sum of the percentages by weight of the four primary alkali elements in the coal ash: sodium, potassium, magnesium and calcium. Figure 12 presents a well-correlated relationship between this sum and the measured dustcake areal density values. Other versions of this expression, adjusted for elemental atomic weight and valence of the individual elements, could also be constructed. Expansion of the database in this paper and field and laboratory investigations into the relative effects of these alkali compounds on ash characteristics should help define the most accurate definition of ash cohesivity.

Though estimations of filtering drag may be similar for ashes with low or high cohesivity, knowledge of the relative contributions of areal density and normalized drag provide information that can aid in the selection of a filtering fabric and cleaning method best suited to the coal being burned. In general, baghouses collecting high-cohesivity ashes should benefit from more energetic cleaning techniques and the use of fabrics with optimum cake release properties. Highly energetic cleaning may not be required to control dustcake weight at baghouses collecting low-cohesivity ashes. In fact, overcleaning of ashes with very low cohesivity may promote penetration of ash through the fabric.

CONCLUSIONS

Within the scope of the database covered in this paper, five conclusions can be drawn from this work:

- Dustcake ash alkalinity and certain chemical constituents in the dustcake ash correlate well with measured areal density and normalized drag values.
- There is good correlation between alkali constituents (sodium, potassium, magnesium and calcium) in coal ash and dustcake ash.
- The measured concentrations of calcium and magnesium in coal ash samples can be used to estimate areal density and normalized drag values, the components of filtering drag.
- Coal ash alkalinity, specifically calcium or magnesium content, is better correlated to baghouse performance than coal sulfur content.
- The empirical relationships presented in this paper provide valuable information for selection of the filtering fabric and cleaning method best suited to a specific coal chemistry.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

ACKNOWLEDGMENTS

The authors wish to thank Michael S. Robinson for his aid in the statistical analyses of the database and for his help in the preparation of this paper.

The data and analyses presented in this paper were funded by EPRI Contract 1129-8.

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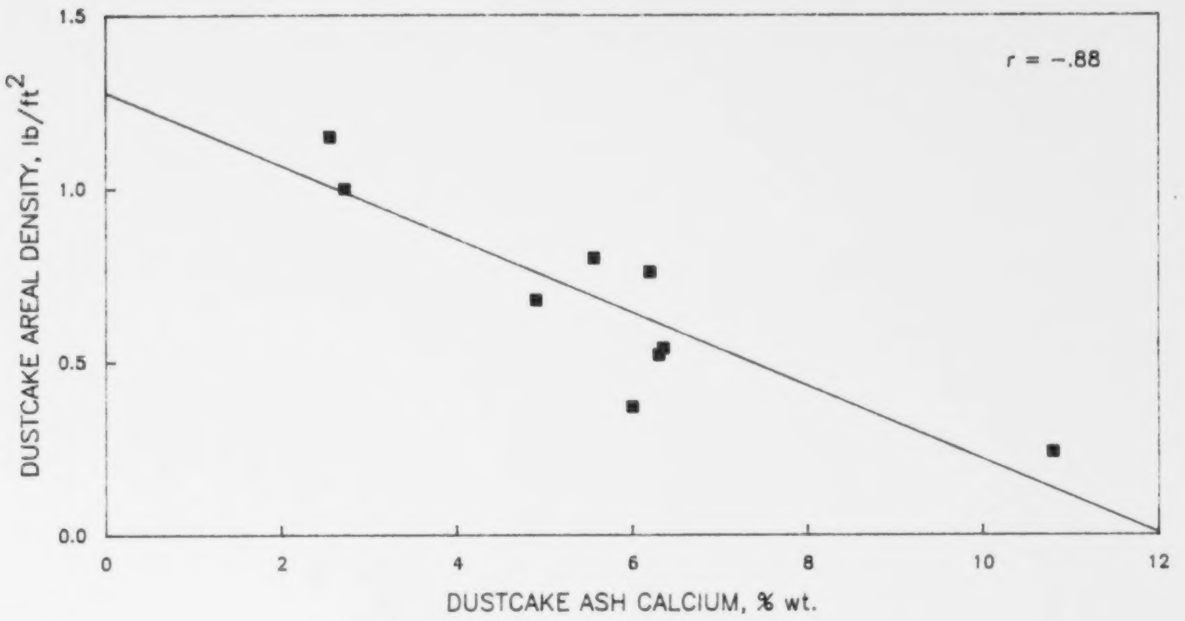

Figure 1. The observed relationship between average dustcake areal density and percentage by weight of calcium in the dustcake ash. The correlation coefficient, r, for a least-squares fit to the data is shown.

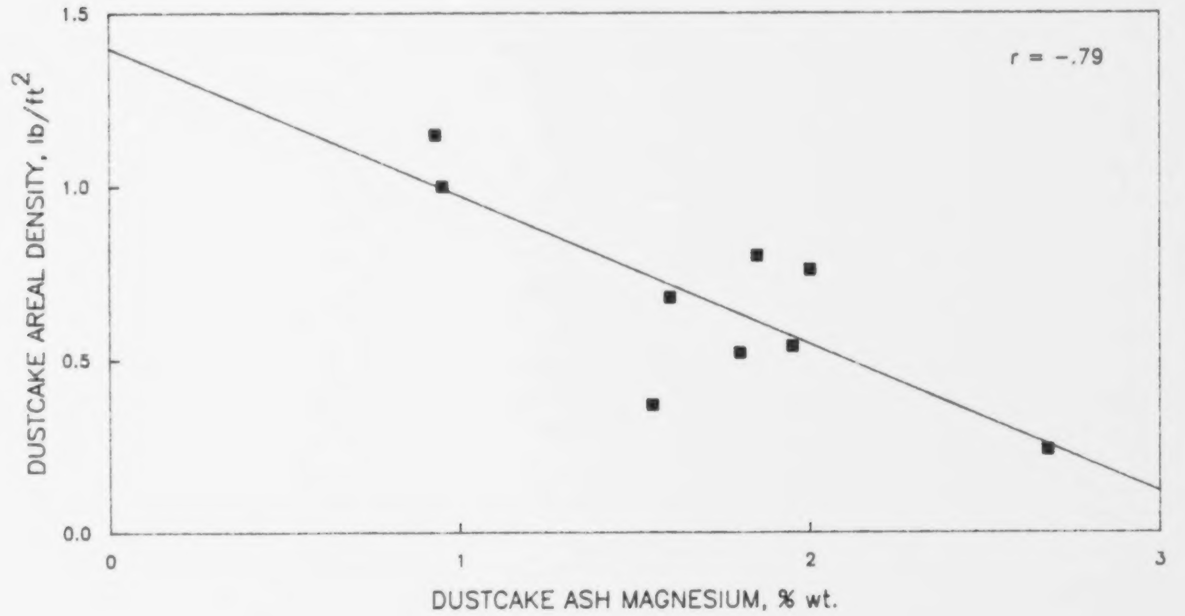

Figure 2. The observed relationship between average dustcake areal density and percentage by weight of magnesium in the dustcake ash. The correlation coefficient, r, for a least-squares fit to the data is shown.

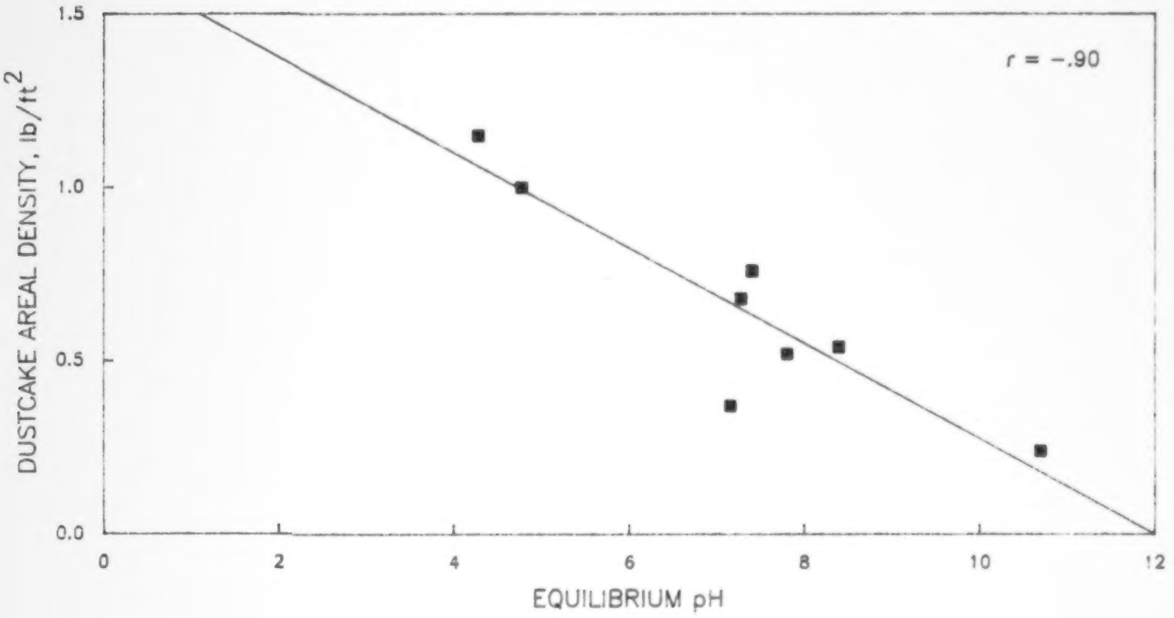

Figure 3. The observed relationship between average dustcake areal density and the measured equilibrium pH of the dustcake ash. The correlation coefficient, r, for a least-squares fit to the data is shown.

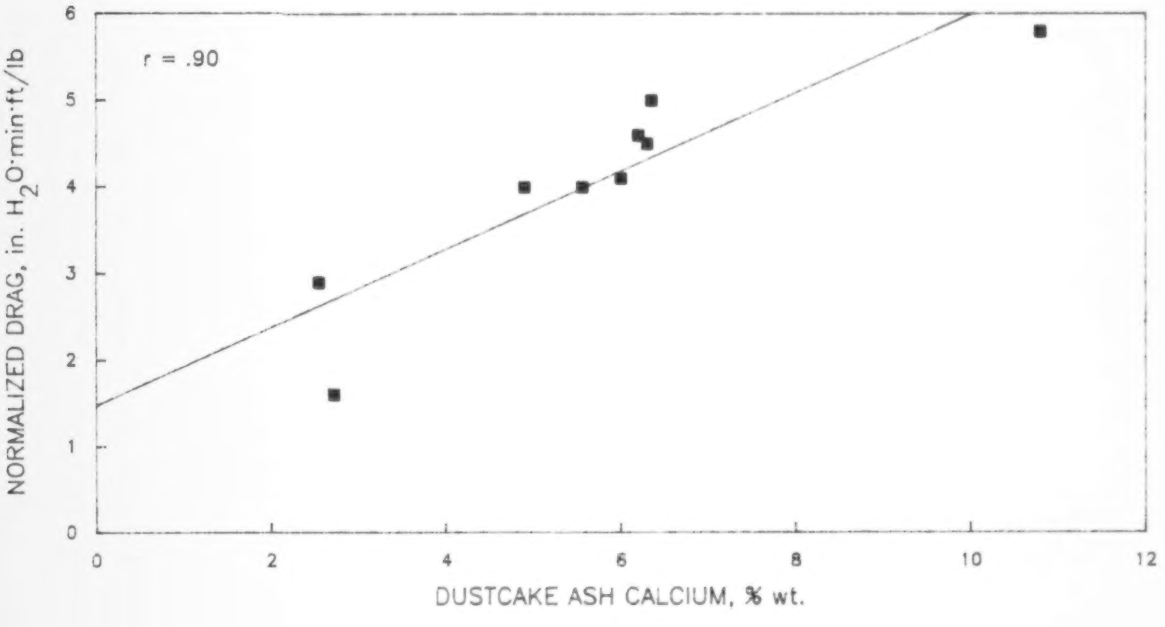

Figure 4. The observed relationship between normalized drag and percentage by weight of calcium in the dustcake ash. The correlation coefficient, r, for a least-squares fit to the data is shown.

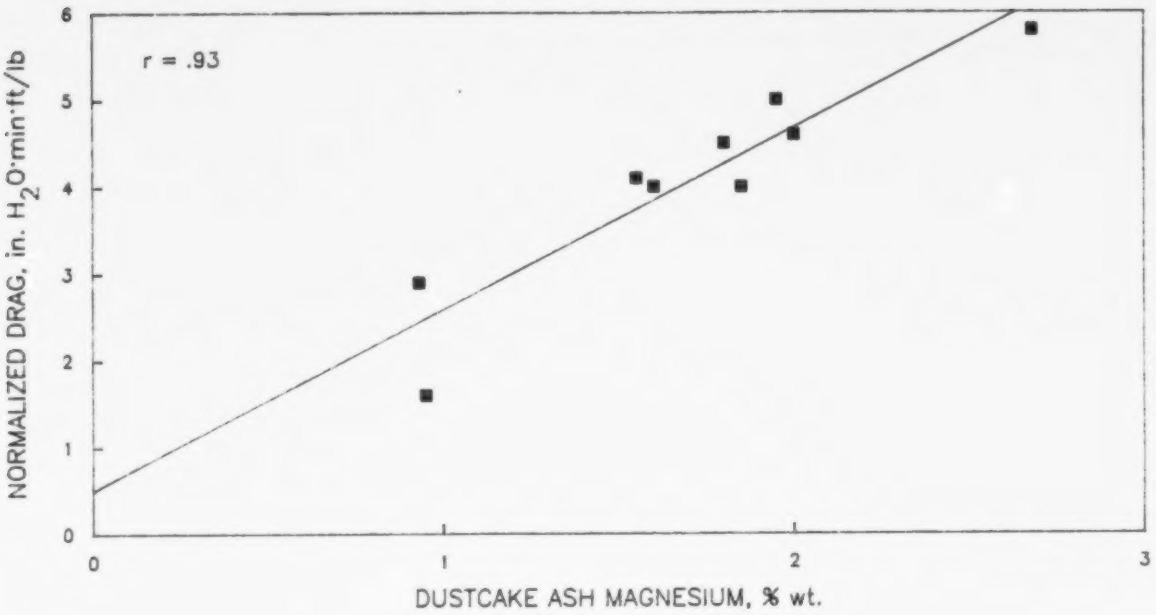

Figure 5. The observed relationship between normalized drag and percentage by weight of magnesium in the dustcake ash. The correlation coefficient, r, for a least-squares fit to the data is shown.

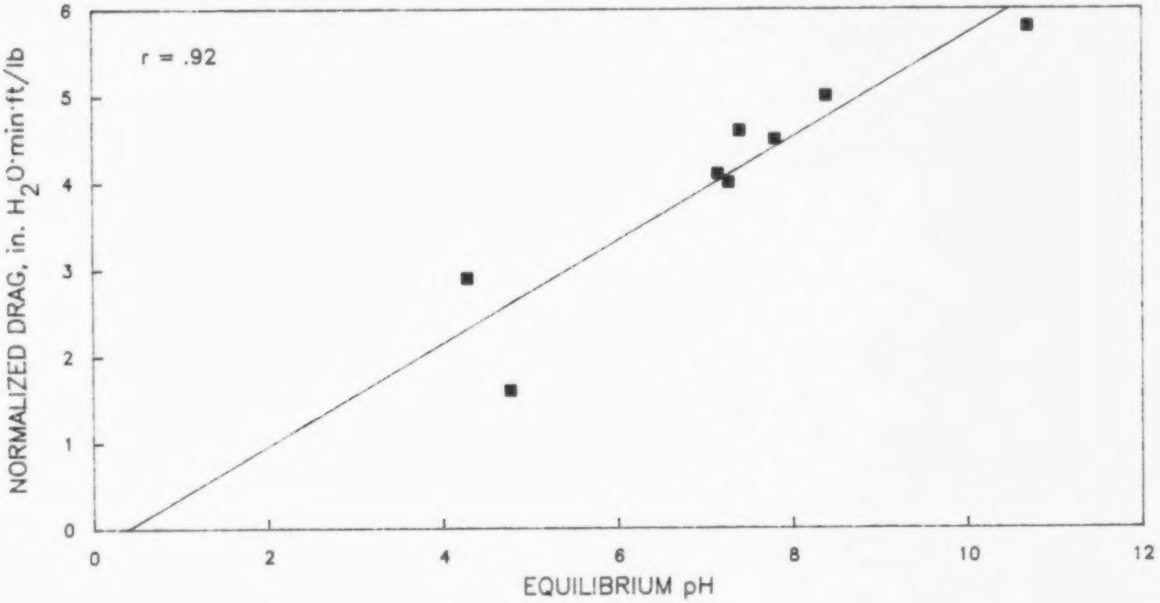

Figure 6. The observed relationship between normalized drag and the measured equilibrium pH of the dustcake ash. The correlation coefficient, r, for a least-squares fit to the data is shown.

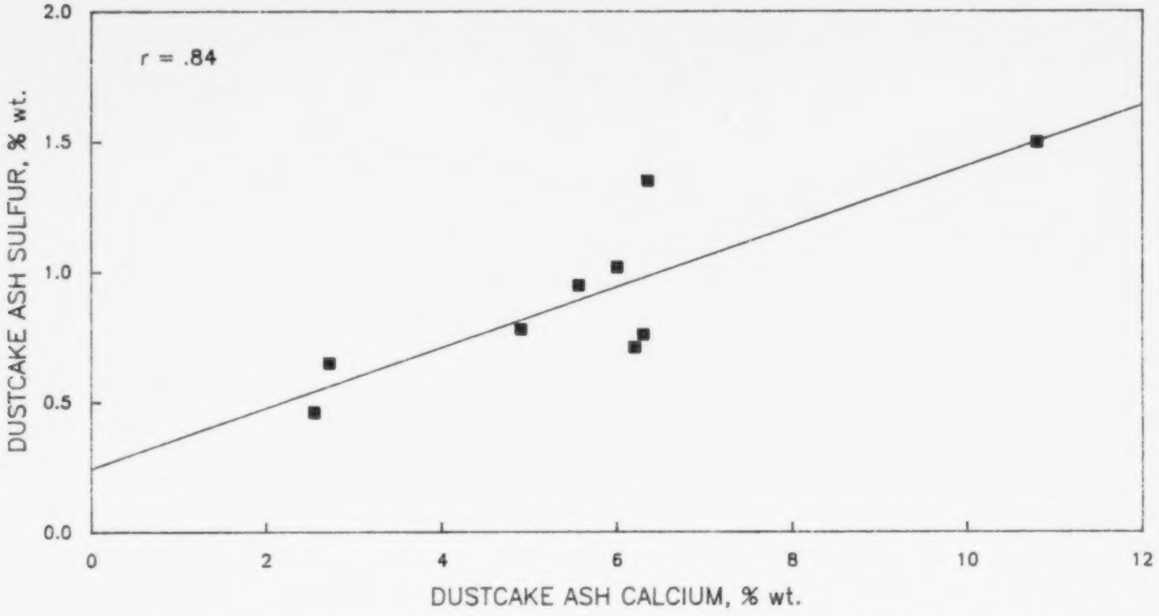

Figure 7. The observed relationship between the percentage by weight of sulfur in the dustcake ash to the percentage by weight of calcium in the dustcake ash. The correlation coefficient, r, for a least-squares fit to the data is shown.

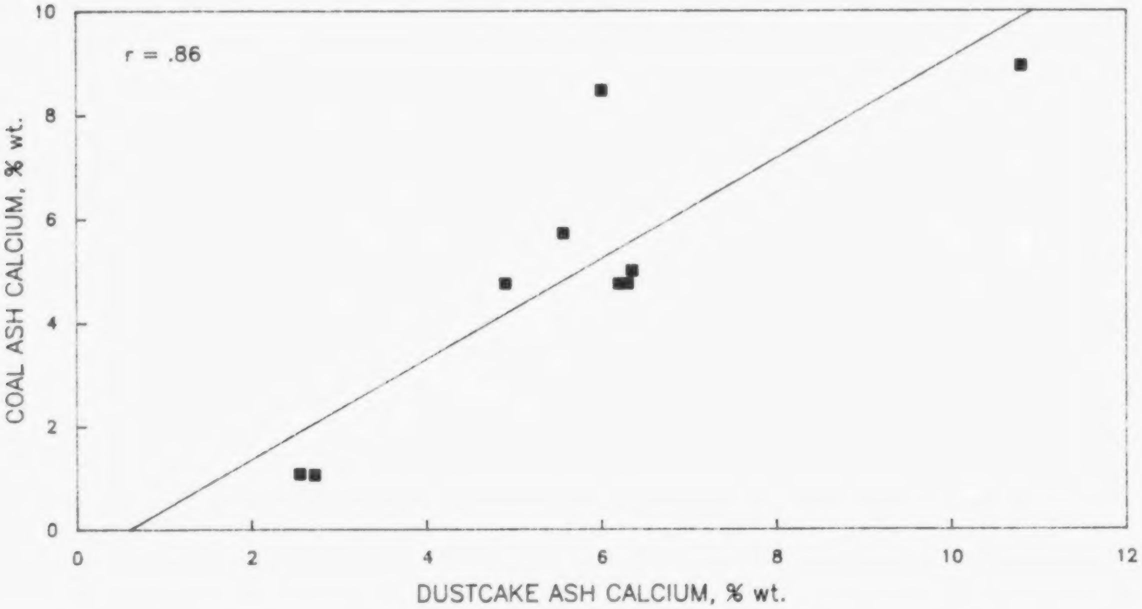

Figure 8. The observed relationship between the percentage by weight of calcium in the coal ash and the percentage by weight of calcium in the dustcake ash. The correlation coefficient, r, for a least-squares fit to the data is shown.

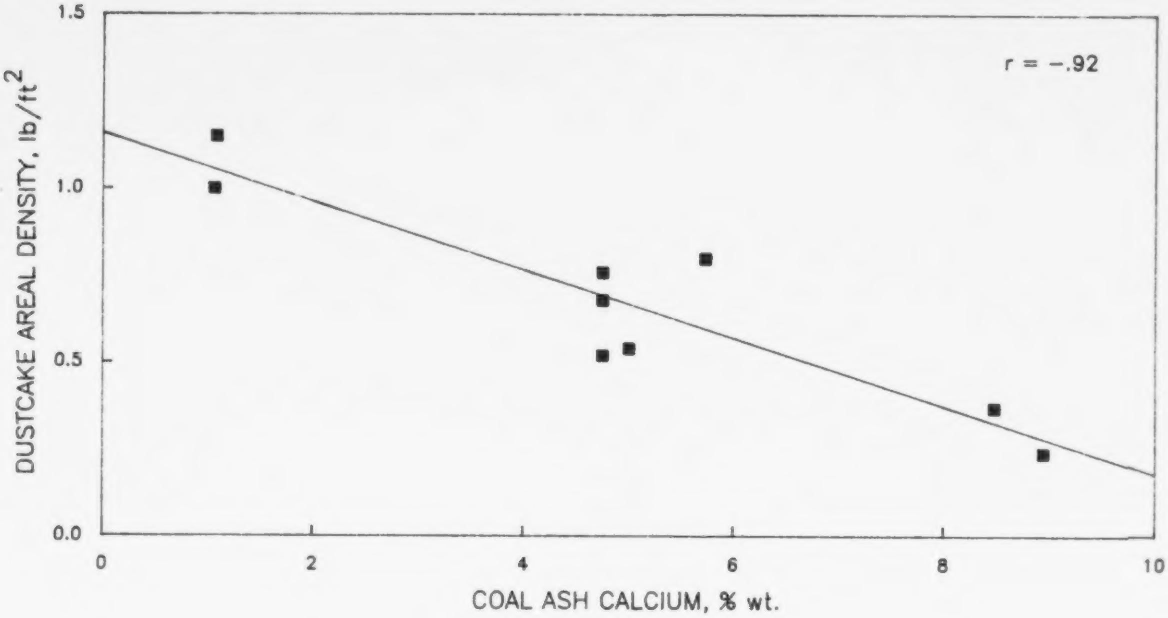

Figure 9. The observed relationship between average dustcake areal density and percentage by weight of calcium in the coal ash. The correlation coefficient, r, for a least-squares fit to the data is shown.

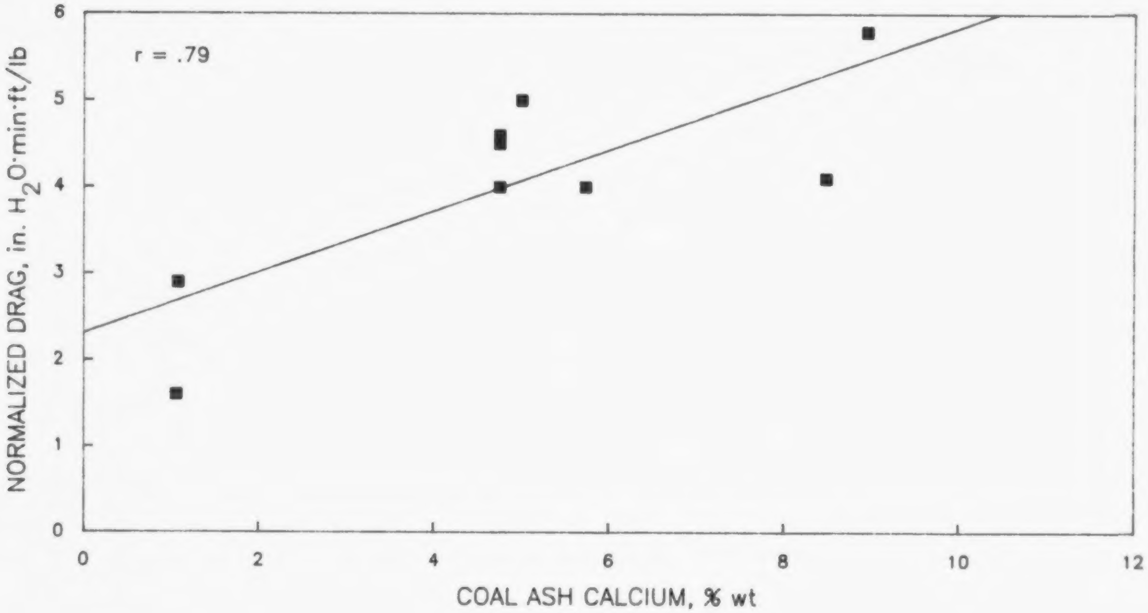

Figure 10. The observed relationship between normalized drag and percentage by weight of calcium in the coal ash. The correlation coefficient, r, for a least-squares fit to the data is shown.

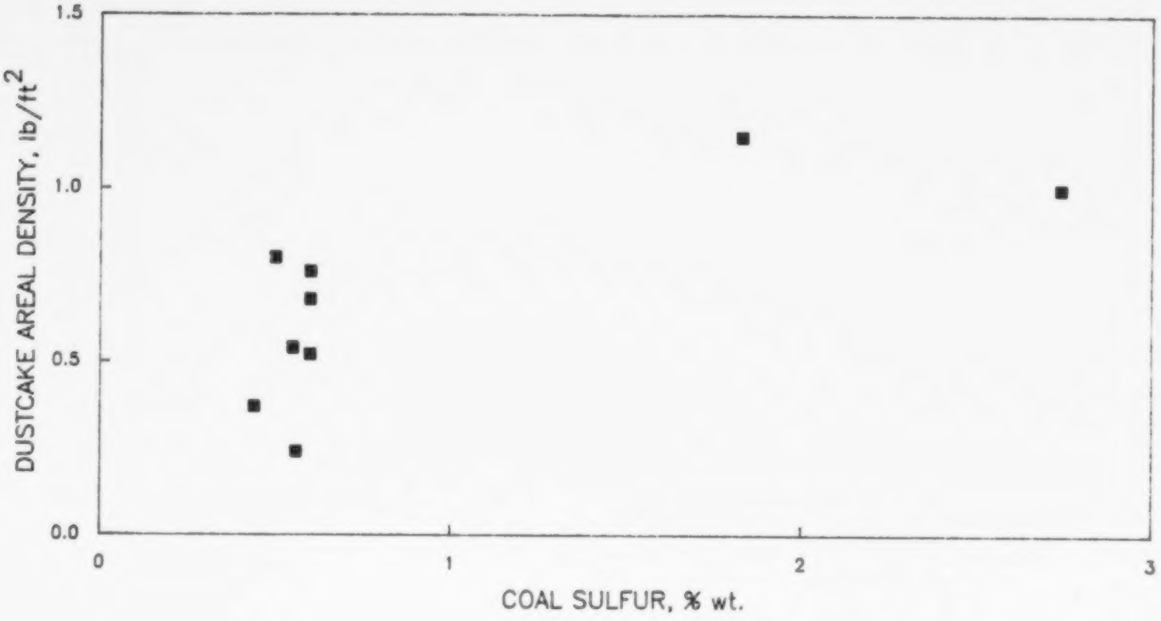

Figure 11. The observed relationship between average dustcake areal density and the percentage by weight of sulfur in the coal samples.

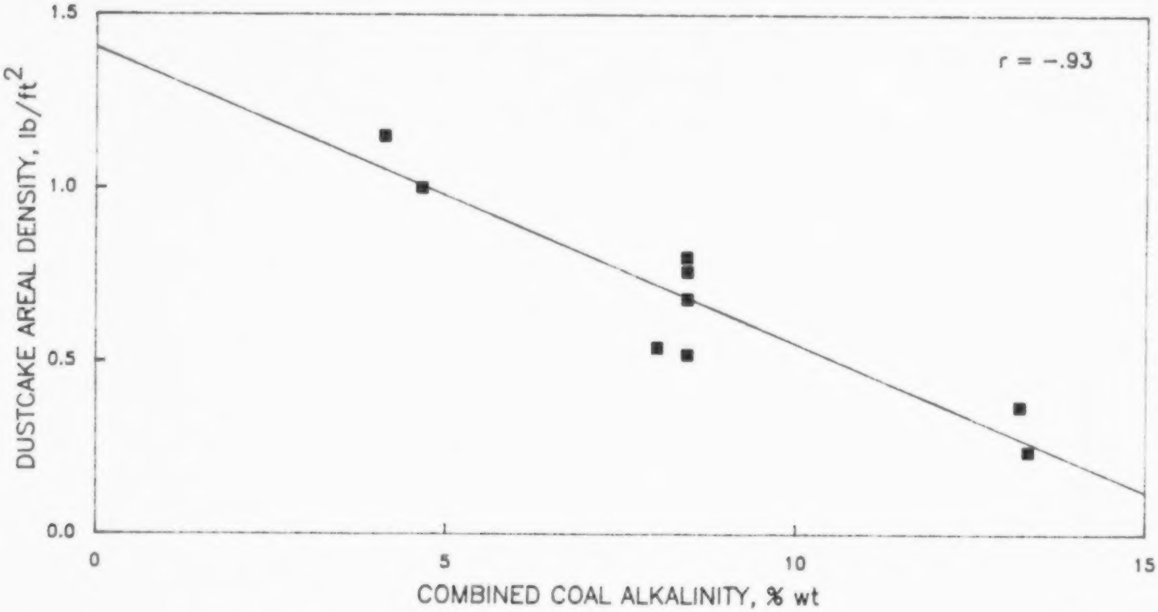

Figure 12. The observed relationship between the total alkalinity of the coal ashes (expressed as the sum of the percentages by weight of sodium, potassium, magnesium, and calcium) and the measured dustcake areal density values. The correlation coefficient, r, for a least-squares fit to the data is shown.

Table 1
BAGHOUSE PERFORMANCE DATA, CHEMICAL CHARACTERISTICS OF DUSTCAKE ASHES, AND COAL CHEMISTRY

BAGHOUSE PERFORMANCE DATA

BAGHOUSE	TUBESHEET ΔP, in. H_2O	AIR-TO-CLOTH RATIO, acfm/ft^2	DRAG, in. H_2O·min/ft	DUSTCAKE AREAL DENSITY, lb/ft^2	NORMALIZED DRAG, in. H_2O·min·ft/lb
SCHOLZ HSFP[a]	2.5	1.6	1.6	1.00	1.6
BRUNNER ISLAND	7.0	2.1	3.4	1.15	2.9
IPP[b]	3.2	2.3	1.4	0.24	5.8
CAMEO	3.5	1.3	2.7	0.54	5.0
ARAPAHOE UNIT 4	3.8	1.6	2.3	0.52	4.5
ARAPAHOE FFPP[c]	5.4	2.0	2.7	0.68	4.0
ARAPAHOE ECOLAIRE	5.2	1.5	3.5	0.76	4.6
CHEROKEE	4.1	1.3	3.2	0.80	4.0
NIXON	2.3	1.5	1.5	0.37	4.1

CHEMICAL CHARACTERISTICS OF DUSTCAKE ASHES, % wt

BAGHOUSE	Na_2O	K_2O	MgO	CaO	Fe_2O_3	Al_2O_3	SiO_2	TiO_2	P_2O_5	SO_3	Li_2O	LOI	SOLUBLE $SO_4^{=}$	EQUILIBRIUM pH*
SCHOLZ HSFP	0.58	2.83	0.95	2.72	21.07	22.43	46.28	1.42	0.55	0.65	0.04	11.10	3.65	4.77
BRUNNER ISLAND	0.35	2.70	0.93	2.55	10.90	29.05	49.50	1.60	0.77	0.46	0.07	6.95	3.55	4.28
IPP	2.12	1.46	2.68	10.80	6.14	19.44	52.90	1.10	0.60	1.50	0.03	0.63	1.92	10.70
CAMEO	0.74	1.20	1.95	6.35	5.00	28.40	50.45	1.70	1.55	1.35	0.03	1.74	1.35	8.39
ARAPAHOE UNIT 4	1.60	1.30	1.80	6.30	5.20	25.70	53.40	0.83	1.30	0.76	0.02	2.60	2.00	7.80
ARAPAHOE FFPP	2.80	1.30	1.60	4.90	4.10	26.40	55.50	1.10	1.00	0.78	0.02	1.80	1.50	7.27
ARAPAHOE ECOLAIRE	1.40	1.30	2.00	6.20	5.70	26.30	52.00	0.92	1.60	0.71	0.02	3.00	1.80	7.40
CHEROKEE	0.57	1.09	1.85	5.56	4.58	28.25	52.08	1.48	1.69	0.95	0.02	1.60	1.07	—
NIXON	2.60	1.10	1.55	6.00	4.80	28.40	50.20	0.96	1.65	1.02	0.03	2.40	2.00	7.15

	COAL ASH CHEMISTRY, % wt										COAL CHEMISTRY, % wt	
BAGHOUSE	Na_2O	K_2O	MgO	CaO	Fe_2O_3	Al_2O_3	SiO_2	TiO_2	P_2O_5	SO_3	SULFUR	ASH
SCHOLZ HSFP	0.48	2.33	0.77	1.06	30.63	22.06	39.06	0.93	0.59	1.23	2.74	9.67
BRUNNER ISLAND	0.40	2.02	0.62	1.08	11.16	23.45	46.13	1.69	—	—	1.83	14.54
IPP	1.23	0.97	2.17	8.95	4.58	15.96	59.25	0.80	0.36	5.15	0.56	9.68
CAMEO	0.73	1.01	1.28	5.00	4.40	25.73	55.69	0.81	1.12	3.74	0.55	8.86
ARAPAHOE UNIT 4	1.23	1.12	1.34	4.75	7.37	24.28	51.51	0.90	1.09	4.82	0.60	6.76
ARAPAHOE FFPP	1.23	1.12	1.34	4.75	7.37	24.28	51.51	0.90	1.09	4.82	0.60	6.76
ARAPAHOE ECOLAIRE	1.23	1.12	1.34	4.75	7.37	24.28	51.51	0.90	1.09	4.82	0.60	6.76
CHEROKEE	0.58	0.93	1.20	5.73	3.37	28.29	53.21	0.84	1.30	4.02	0.50	8.14
NIXON	2.64	0.71	1.38	8.48	6.18	21.90	44.74	0.81	1.78	10.47	0.44	4.45

[a] EPRI High-Sulfur Fabric Filter Pilot Plant

[b] Intermountain Power Project Unit 1

[c] EPRI Fabric Filter Pilot Plant

* Dimensionless

LIFETIME BAG HOUSE COSTS

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ABSTRACT

Lifetime bag house cost calculations, differentiating between investment and operating cost elements, were developed to reflect current perceptions of costs and then translated for use on a spread sheet computer program. With the help of this program, cost sensitivity to variables was explored. The analysis quantifies the high dependence of lifetime costs on factors which influence investment costs, such as gas/cloth ratio and construction costs, while determining that costs associated with operating labor, filter bag life, and pressure drops had much less influence. The effect of changing a single calculation parameter by 50%, ranged from a high of 32% for "bag house investment" to less than 1% for "plant labor costs".

INTRODUCTION

Lifetime bag house costs, as differentiated from out of pocket annual operating costs, not only include the usual cost items such as bag replacement, maintenance, labor, operating utilities, and the like, but also the costs associated with the moneys required initially to purchase the bag house. The true cost should reflect the bag house investment, even though it generally does not produce goods and is only there as a required item to permit one to operate an earnings producing process. The value of the bag house investment should be treated in the same manner as if the funds used to purchase the bag house were allocated for an income producing activity, and the cost of the bag house to the owner should reflect that in terms of: "If I had this sum to put into an income producing activity, what would my required return on this investment have to be for me to invest it?"

With this consideration in mind, we will review the effect of various parameters used to calculate annual bag house costs on the annual bottom line. The parameters which influence costs can be divided into two classes: Investment costs and operating costs. Items considered in this study, and their grouping are shown in Table 1.

As one can readily see, the investment items reflect the costs associated with putting the entire unit in place and preparing it for operations, while the operating costs are the costs associated with keeping it going once it is in place.

These groupings, in terms of what inputs influence their quantity, and how these quantities are derived, will now be discussed.

INVESTMENT COSTS

The BAG HOUSE investment is determined by the size of the bag house, which in turn, is determined by the amount of gas to be cleaned, the gas/cloth ratio for which it is designed, and the associated hardware, such as cages, fans, and valves. The quantitative aspects of this will be discussed shortly. (Table 2)

The major factors influencing the bag house costs are: The size of the bag house, as influenced by the gas/cloth ratio of the design, and the cost of the money for purchasing the bag house. This latter cost is often expressed as "Required Pretax Earnings on Investment" or a similar term, which denotes the minimum return on investment required by the source of the funds. It could be an interest charge by a lender, or an internal corporate requirement and is considered as an operating cost.

The BAG COSTS are an additional investment item, which is treated as a contributor to the initial purchase cost. As will be seen later, the cost of replacing the bags as they wear out is treated as an operating cost. Factors influencing bag costs are shown in Table 3.

The bag cost components are: materials costs, materials yields, bag size, number of bags, and special bag hardware.

Finally, investment costs usually include the cost of a certain number of spare bags which are supplied initially to compensate for installation and start up problems. This quantity is generally 3 percent of the required number of installed bags.

To summarize, the investment costs are comprised of the sum of the bag house cost, bag cost, and spare bags.

OPERATING COSTS

Operating Costs are those costs which recur annually, and are required to keep the bag house, once installed, in operation.

Some operating costs are set by convention, such as: Table 4.

Annual depreciation and maintenance values are generally expressed as a percentage of the Investment cost, and reflect actual operating experience, as established by local experience, and by the estimated useful life of the investment.

The costs associated with replacing worn bags are determined by the bag cost and the replacement frequency, and include not only the cost of replacement bags, but also the cost of removing the old bags and replacing them with new ones. (Table 5)

An estimate of the bag life will determine the bag replacement rate, while the bag cost determines the funds required for replacement bags. The costs of changing the bags is determined by the local plant labor cost (full cost).

The final cost elements to be considered are those associated with the processes of pulling the gases through the bag house and the costs of cleaning the bags while in service (Table 6).

Elements which affect these costs are: Pressure drop across the tube sheet, cost of local electricity, and average cleaning air flow.

INPUT DATA

Calculation of the various cost contributions require a substantial amount of data. The Data required, and the units are listed in the next slide (Table 7).

CALCULATED DATA

The calculated data, derived from the input data, is shown in (Table 8), as are the units in which they are reported, and the method used to derive their value.

Now to return to the remaining, missing item, namely the calculation of a unit cost value for the bag house. To do this we used a recent article on this subject[1] by J. H. Turner and J. D. McKenna. In that article it is stated that the cost of a modular pulse-jet filter with insulation is arrived at by using the formulae shown in: (Table 9).

ANNUAL COST CALCULATIONS

With the data now in hand, we are able to proceed with a calculation of the annual costs. This was done by translating the methodology onto a spread sheet, and calculating the costs. The derived values were reported in two groupings: investment costs and operating costs. Both of these costs were then summed, and expressed in costs per annum per rated ACFM capacity. A detail of the calculated costs for the case discussed (which we call the "base case"), is shown in (Table 10)

Alternate cases, in which each one of the input variables discussed were changed by 50%, were calculated to assess the sensitivity of the annual costs to such a change. As a result of this phase of the study we obtained a read out on the sensitivity of the annual coat to changes. The results obtained are given in Table 11.

Looking at the results from the typically cited power plant operator's point of view, we immediately see the importance of the bag cost and pressure drop variables on the annual operating cost (Table 12).

However, reviewing the effect of changing variables on the entire costs clearly shows that the investment cost in the bag house, and the parameters which determine that cost, are far more important in establishing the annual costs, than are operating parameters such as pressure drop power costs, and bag costs. (Table 13)

Thus, a clear case is made that the design of a bag house, so as to minimize initial investment, by (for example) proper utilization of the full potential of filter media, is of prime importance. Only in retrofit installations, where the bag house investment costs are already fixed, do the variables that we normally worry so much about take on an aspect of primary importance.

REFERENCES

1. J. H. Turner, A. S. Viner, R. E. Jenkins, W. M. Vatavuk and J. D. McKenna: "Sizing and Costing of Fabric Filters"; JAPCA Vol. 37, No. 9 September 1987.

The work described in this paper was not funded by the U.S. Environmental Protection Agency. The contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

TABLE 1. PARAMETERS INFLUENCING BAG HOUSE COSTS AND THEIR CLASSIFICATION

Investment Costs	Operating Costs
Bag House Investment	Annual Depreciation
Initial Bag Investment	Maintenance
Spare Bag Investment	Replacement Bags
	Bag Change Costs
	Fan Power Cost
	Cleaning Air Cost

TABLE 2. PARAMETERS AFFECTING BAG HOUSE INVESTMENT COSTS

Bag House Unit Cost	$ / sq. ft. Filter Area
Bag House Capacity	cu. ft. / min.
Gas to Cloth Ratio	cu. ft. / sq. ft. / min.

TABLE 3. PARAMETERS AFFECTING BAG INVESTMENT COSTS

Bag Material Cost	$ / sq. yd.
Bag Area	Bag Dimensions
Fabric - Bag Conversion	% Yield
Special Bag Hardware	$ / bag

TABLE 4. PARAMETERS AFFECTING OPERATING COSTS SET COSTS

Annual Depreciation	% of Total Investment
Maintenance	% of Bag House Investment

TABLE 5. PARAMETERS AFFECTING BAG REPLACEMENT COSTS

Replacement Bags	Bag Life in years Bag Cost $ / bag
Bag Change Costs	Change Rate bags / hour Plant Labor Rate $ / hour

TABLE 6. PARAMETERS AFFECTING AIR FLOW COSTS

Pressure Drop	inches of water
Power Costs	$ / KWH
Cleaning Air Flow Rate	cu.ft. / min.

TABLE 7. INPUT DATA DESCRIPTION AND UNITS

Investment Costs	
Bag House Capacity	- cu. ft./min.
Air/Cloth Ratio	- cu.ft./sq.ft./min.
Bag Area	- sq.ft.
Bag Material Cost	- $/sq.yd.
Bag Preparation Yield	- %
Bag Life	- years
Bag House Unit Investment Cost	- $/sq.ft.
Req'd. Pretax Earning on Inv.	- %
Operating Costs	
Operating Pressure	- in. H2O
Power Unit Cost	- $/cu.ft./min./in. H2O
Cleaning Air Cost	- $/cu.ft./min.
Depreciation Rate	- % of Investment
Maintenance Cost Rate	- % of Investment
Plant Labor Cost	- $/hr.
Power Cost	- $/KWH
Average Cleaning Air Flow Rate	- cu.ft./min.
Bag Change Rate	- # bags/man/hr.

TABLE 8. CALCULATED VALUES

Investment Costs	Units	Method used to Derive Value
Bag Cost	- $/bag	- bag area x Yield x Mat.Cost
Filter Surface Area	- sq.ft.	- Capacity / A/C Ratio
# of Bags	- #	- Filter Surface Area/Bag Area
Installed Bag Cost	- $	- # bags x Bag Cost
Bag House Unit Cost	- $/sq.ft.	- (see next Slide)
Bag House Cost	- $	
Spare Bag Costs	- $	- 3 % of Installed Bag Cost
Operating Costs		
Annual Depreciation	- $/yr	- Invstmnt. x Depr. Rate
Maintenance	- $/yr.	- Invstmnt. x Maint. Cst. Rate
Replacement Bag Cost	- $/yr.	- Instld. bag cst./bag life
Bag Change Cost Cost	- $/yr.	- # bags/baglife/bagchgrt. x Labor
Fan Power Unit Cost	- $	- $/Press.Drop/cu.ft./min.
Fan Power Cost	- $/yr.	- FanPwrUnCst x Capacity
Cleaning Air Cost	- $/yr.	- Cln.AirUnCst x Flow Rate

TABLE 9. CALCULATION OF BAG HOUSE UNIT COST

Formulae (1)

Size:

$ = (47,220 + 7.323(sq.ft.)) + (3,000 + 2.079(sq.ft.))

Cages:

$ = (10.024 + 0.163(sq.ft.))per cage = $ 11.73 per cage

Bag House

Capacity = 60,000 acu.ft./min.
Air/Cloth Ratio = 5.0 cu.ft./sq.ft./min.
Cloth Area = 60,000 / 5.0 = 12,000 sq.ft.
Bag Area = 1.16 sq. yd./bag
Number of bags = 12,000/1.16 x 9 = 1150

Calculation

Bag House:

$ 50,220 + 9.402 x 12,000 = $ 163,044

Cages:

$ 11.73 x 1150 = $ 13,490

TOTAL (Flange-flange) = $ 176,534
TOTAL (Installed) (2.9x) = $ 511,950
UNIT COST = $ 42.66 / sq. ft. filter area

TABLE 10. CALCULATED COSTS
Base Case

A. Investment

Bag House = $ 510,000

Bags (+ Spares) = $ 74,790

Total = $ 584,790

Unit Investment = $ 9.75 / ACF / yr.

Annual Cost (9.75 x 0.15) = $ 1.46 / ACF /

B. Operating Costs

Depreciation = $ 46,783

Maintenance = $ 20,400

Replacement Bags = $ 24,204

Bag Change Cost = $ 3,439

Fan Power Cost = $ 32,370

Cleaning Air Cost = $ 1,000

Total = $ 128,197

Annual Operating Cost = $ 2.14 / ACF /

TOTAL ANNUAL COST = $ 3.60 / ACF

TABLE 11. SENSITIVITY ANALYSIS
EFFECT OF CHANGING VARIABLES BY 50%

Variable	Value		Annual Cost ($/ACFM)		
	From	To	Investment	Operating	Total
Base Case	---	----	1.46	2.14	3.60
A/C Ratio	5.0	7.5	0.97	1.61	2.58
Bag Cost	63.33	95.01	1.56	2.39	3.94
Bag Life	3.0	4.5	1.46	1.98	3.45
Press. Drop	5.0	7.5	1.46	2.41	3.87
Bg. Hs. Un. Cst.	42.50	63.75	2.10	2.65	4.75
Plnt. Lbr.	27.00	44.50	1.46	2.17	3.63
ROI	15	10	0.97	2.14	3.11
Depr. Rate	8	12	1.46	2.53	3.99
Maint. Cst.	4	6	1.46	2.31	3.77
Bag Length	8	12	1.46	2.12	3.58
Power Cost	0.05	0.08	1.46	2.46	3.92
Clng.AirCST	0.33	0.50	1.46	2.14	3.60

TABLE 12. ANNUAL BAG HOUSE OPERATING COSTS

Variable	Annual Cost	% Change
Base Case	$ 2.14	---
Bag Cost	$ 2.39	11.7
Pressure Drop	$ 2.41	12.6
Bag Life	$ 1.98	7.5
Plnt. Lab. Cst.	$ 2.17	1.4

TABLE 13. GROUPING OF BAG HOUSE COSTS

Cost Parameter	Influence Range	Type of Cost
	> 25 %	
Bag House Unit Cost	31.9	Investment
Air-Cloth Ratio	28.3	Investment
	25 - 15 %	
	None	
	15 - 10 %	
Reqd. Ret. on Investment	13.6	Operating
Depreciation Rate	10.8	Operating
	10 - 5 %	
Bag Cost	9.4	Inv. + Op.
Power Cost	8.9	Operating
Pressure Drop Cost	7.5	Operating
	< 5 %	
Maintenance Cost	4.7	Operating
Bag Life	4.2	Operating
Plant Labor Cost	0.8	Operating
Bag Length	0.6	Operating

RECOMMENDATIONS FOR A BAGHOUSE DESIGN CHANGE AND FOR SOME NEW TEST PROCEDURES

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ABSTRACT

Conventional reverse air (RA) cleaning systems recirculate very fine particulate from compartments being cleaned to the downstream compartments in service. The RA recirculation has several detrimental consequences; it increases pressure loss by reducing particle size distribution in the cake; it causes unequal compartment performance; it increases residual cake pressure loss and it adds cake to the clean side of the bags. Recapturing and recirculating fine particles in a conventional RA system is not necessary.

This paper describes a system that eliminates these adverse characteristics of conventional RA systems.

In another paper (# 2B4) the author discusses some of the things we do not know about fabric filtration and suggests some subjects for further research. This paper describes some new test procedures that can be used to provide data that have heretofore not been available and that can facilitate the recommended research.

INTRODUCTION

This paper pertains to the large custom-designed baghouses that are used to remove fly ash from the flue gas of coal-fired electric utility power plants. In particular, it pertains to bottom inlet, suction-type baghouses that are cleaned by conventional reverse air (RA) systems.

RECOMMENDED DESIGN CHANGE

One of the reasons for proposing a design change is because of the author's premise that fine particles accumulate in the top of a bottom inlet bag and that they are not adequately removed by conventional RA cleaning. Another reason is the fact that there are some baghouse performance anomalies that have not been adequately explained. The premise refutation and its justification are as follows.

PREMISE REFUTATION

The author's premise has been refuted by several investigators who used questionable techniques to arrive at their conclusion that fines do not accumulate in the top of a bottom inlet bag. These investigators isolated compartments in operating baghouses and inspected the off-line bags to see if there was an accumulation of fines in the top. As could be expected, they found that there was not. The author believes that these investigations are inconclusive.

When a compartment is isolated there is no pressure differential across the bags. Some cake, held in place by on-line pressure differential, may fall off from any place in the bag, including the top. Assume that the access doors were opened while the compartment was not under vacuum and, after the doors were opened, that the inlet damper was "cracked" open to provide some suction to draw outdoor air through the compartment to purge and cool it. Assume some length of time, say an hour, to purge and cool so that the compartment could be entered. During that time cool air flowed in the reverse direction through and downwards in the bags, as in reverse air cleaning. During reverse air cleaning, a compartment will have reverse flow for a minute or less per cycle. During these investigations the off-line compartments probably experienced reverse flow for an hour or more.

On entering the compartment, the investigators, moving around on the tubesheet and on the upper walkways, undoubtedly caused some agitation of the bags. Even minor agitation can cause cake release. In order to find out what was inside the bags, the investigators had to cut swatches with the bags in place or remove them in order to cut them open. This further disturbance of the bags most likely caused some additional cake release. Thus, prior to examination, the test bags were subjected to reverse air cleaning, "shaking", and manual disturbance. Moreover, in service, there will be convection currents and stack effect at operating temperature that will tend to hold small particles in the top of the bag, even during reverse air cleaning. By the time these test bags were examined, these influences were negligible.

The author believes that what remained in those test bags at the time they were examined was different from what was in the bags when they were in service. These investigations do not invalidate the premise. You cannot do an autopsy on a dead bag to determine what it was doing when it was alive.

PREMISE JUSTIFICATION

Theory supports the premise as follows.

The gas velocity entering the bottom of a bag can carry particles up to some maximum size. The volume of gas rising in a bag diminishes as it rises because of flow through the bag. Because of the constant cross-sectional area of the bag and the diminishing gas volume the velocity in the bag decreases from the entrance velocity at the bottom to zero velocity at the top of the bag. The ability to carry particles decreases as gas velocity decreases. This logic tells us that the particle size distribution entering the bottom of the bag will change as the gas flows up in the bag to a size distribution containing mostly very fine particles at the top of the bag.

Reference (1) provides a formula relating particle size and carrying ("floating") velocity. The formula is as follows:

$$V_f = \sqrt{\frac{4}{3}\ \frac{g}{f_D}\ \frac{\delta p}{\delta a}\ d_p} \qquad 1$$

V_f = floating velocity, ft/sec
g = acceleration of gravity, ft/sec^2
f_D = drag coefficient, dimensionless
δp = average bulk density of particle, lb/ft^3
δa = density of gas (air), lb/ft^3
d_p = particle diameter, ft

Assume that the particles are spheres and the f_D is therefore 0.5. Also assume that δ_p is 100 lb/ft^3 and that our gas is air at 300°F with $\delta a = 0.0523$ lb/ft^3. One foot = 304,800 microns. For V_f in ft/min and d_p in microns our formula is thus:

$$V_f = 60\sqrt{\frac{4 \times 32.2 \times 100 \times d_p}{3 \times 0.5 \times 0.0523 \times 304{,}800}} = 44\sqrt{d_p} \qquad 2$$

Assume a bottom inlet bag 30 feet long with a nominal diameter of one foot and without anti-collapse rings. With these assumptions we can calculate the gas velocity entering the bottom of the bag for a range of cloth ratios. That velocity will be zero at the top of the bag. If we assume uniform flow rate through the cloth over the full length of the bag then the decrease in the vertical velocity will be linear. Five feet down from the top of the bag the vertical velocity will be one sixth of the bag inlet velocity.

Using Equation 2 we can calculate the maximum particle size that will be carried to the top five feet of the bag for each of the cloth ratios. The results of these calculations are shown in Figure 1. This theoretical analysis supports the premise.

LOGIC SUPPORTS THE PREMISE AS FOLLOWS

Baghouse pressure loss increases exponentially as particle size decreases. We should, therefore, avoid anything that causes a shift in particle size toward smaller sizes and also avoid anything that causes an increase in the proportion of fines in the grain loading.

When the dirty gas enters a compartment hopper, some of the larger particles will drop out due to the large, abrupt decrease in gas velocity. Hopper fallout causes a decrease in the median particle size entering the bag.

With a conventional reverse air cleaning system, the reverse flow is returned to the inlet duct. The duct connecting the inlet duct to the hopper is usually designed for a gas velocity during on-line flow in the range of 2000 to 2400 fpm. This velocity is adequate to carry the particulate into the hopper because the flow is usually downwards from the bottom of the inlet duct and because the duct is relatively short. The reverse air cloth ratio is usually less than the on-line cloth ratio and the gas velocity in the connecting duct during reverse flow may be in the 1500 to 1800 fpm range. As described in Paper # 2B4 we are collecting astronomical populations of very small particles. During RA flow we can expect that the larger particles and any agglomerated cake will fall into the hopper and that the RA flow will carry a very large population of very small particles to the inlet duct.

Because of hopper fallout and RA flow recirculation we can expect that the median particle size entering the bags will be significantly smaller than the median size entering the baghouse. In addition to those two things, which aggravate our problems, we are increasing the population of particles that can reach the top of the bags according to Figure 1.

Reference (2) includes a number of citations confirming the logic that small particles accumulate in the top of bottom inlet bags.

We have theory and logic to support the premise and some inconclusive tests to refute the premise. We also have some indirect evidence that supports the premise and that provides explanations for some baghouse performance phenomena that seem to be contrary to logic. We can now consider these unexplained baghouse performance anomalies in relation to the premise.

THE ANOMALIES

An early claim for baghouses, especially when compared to electrostatic precipitators, was that baghouses are relatively insensitive to particle size distribution and to inlet grain loading. Another early claim was that if small particles and/or high inlet grain loading caused unacceptable pressure loss, all that was necessary to correct the problem was to increase cleaning frequency and/or RA flow rate. We now know that this claim is not always valid.

Reference (3) explains that baghouses are sensitive to particle size distribution and that particle size distribution is dependent upon some unquantified chemical and physical properties of coal and fly ash. At this time we do not have an acceptable correlation between coal/ash properties and baghouse pressure loss. There does seem to be agreement among those knowledgeable in the art that design face velocity (cloth ratio) should be reduced as particle size diminishes and as the proportion of small particles in the inlet grain loading increases.

Reference (4) reports that increasing cleaning frequency may increase pressure loss. This anomaly has not been explained. Reference (5) confirms that the converse is sometimes true; reducing cleaning frequency may reduce pressure loss.

EXPLAINING THE ANOMALIES

One anomaly can be explained, granting the author's premise, as follows:

Assume a ten-compartment baghouse (the number of compartments does not matter as long as it is four or greater) with two rows of five compartments and with one row on each side of the inlet duct. Number the compartments 1,3,5,7 and 9 on one side and 2,4,6,8 and 10 on the other side of the inlet duct. Assume that the dirty gas inlet to the inlet duct is upstream of compartments 1 and 2.

Assume an inlet duct velocity of 2000 to 4000 fpm or, typically about 3600 fpm, and that the inlet duct is tapered to achieve substantially constant duct velocity.

Assume that the reverse air flow from a compartment hopper back to the inlet duct carries particulate and that this particulate is all in the very small size range.

Assume that when compartments 1 and 2 are cleaned that their RA flow goes to compartments 3 through 10. When compartments 3 and 4 are cleaned, their RA flow goes to compartments 5 through 10. The RA flow from compartments 5 through 8 also goes downstream. When compartment 9 is cleaned, its RA flow goes to compartment 10 and vice versa. These assumptions are based on a judgment that the RA flow leaving a compartment hopper will not flow upstream against a duct velocity of about 40 miles per hour and that it will not be able to cross the inlet duct to enter a hopper directly opposite when there are other hoppers downstream. This analysis is also described in Reference (3).

If these assumptions are correct, there will be a small amount of fines in the tops of the bags in compartments 1 and 2, an increasing amount in the bags in each of the pairs of downstream compartments, and a very high amount in the tops of the bags in compartments 9 and 10.

To illustrate this point, assume a reverse flow of 60,000 ACFM and a reverse flow grain loading of 4.2 gr/ACF and that the reverse flow is for 1 minute per compartment per cycle. The reverse flow will then carry 36 pounds per compartment per cycle to the inlet duct. If we start by cleaning compartment No.1, and, if the fly ash goes downstream only, we then have a distribution of reverse flow to the on-line compartments as shown in Table 1.

The large accumulation of fines in the tops of the bags in compartments 9 and 10 should not be a problem if reverse air cleaning is equally effective for small and large accumulations of fines in the tops of the bags. If there were no problem, we could expect all compartments to collect equal amounts. There may be a problem because the greatest collection is frequently in the low-numbered compartments and the least collection in the high-numbered compartments, especially the two at the end of the inlet duct. This unequal collection may be because there is only a small accumulation of fines in the tops of the bags at the inlet end and an increasing amount of fines in the tops of the bags approaching the dead end of the inlet duct. If the accumulation of fines causes the bags to go blind at the top, then there will be a progressive loss of filtering area in each successive pair of compartments that will be small at the inlet end and which may be very significant for the compartments at the end of the inlet duct.

If these assumptions are correct, we may be asking the RA system to clean easily-cleaned compartments at the inlet and very difficult compartments at the end of the inlet duct. RA cleaning uses two mechanisms; bag deformation and backwash. These mechanisms may achieve different results in different situations.

Assume that prior to RA cleaning in compartments 9 and 10 that there is a large accumulation of fines in the tops of the bags in these compartments. At the start of RA cleaning, the tension in the warp fibers will be the smallest at the bottom of the bags due to the weight of cake on the bags and on the deflection of the support springs. On reversing the pressure differential across the bags, the greatest deformation will take place at the bottom where there is the least warp tension. Since bag shape change is the principle mechanism for breaking the cake loose, it follows that the best cleaning will occur at the bottom and that it will occur there first. The bottom then becomes the path of least resistance and we can expect that the greatest RA flow will go through the bottom. We can also expect that there will be the least change in bag shape at the top because that is where warp tension is always the greatest and where reverse flow is the smallest.

During reverse air cleaning, some of the fines in the top of a bag will stick to the bag, some will agglomerate and fall to the hopper and some will become airborne and float in the bag and the upper part of the hopper. The floaters will return to the bag surface when it goes back on line. The floating component may be very large in the compartments at the end of the inlet duct. So large, that together with the fines that stick to the bag, they may cause a large area, especially at the top of the bag, to go blind. This reduction in cloth area would cause an increase in cloth ratio and pressure loss for the whole baghouse. The fines that stick to the top of the bag and the floaters may be significant causes for high residual cake pressure losses.

Logic tells us that increasing cleaning frequency should reduce pressure loss. The greater the cleaning frequency, the cleaner the bags, and the lower the pressure loss. In fact this is not necessarily so. This anomaly has not been explained. If we increase cleaning frequency, we also increase the amount of RA fines going back to the inlet duct and thus accelerate the rate of accumulation of fines in the tail-end compartments. This may explain why some baghouses have experienced an increase in pressure loss when cleaning frequency was increased and a lower pressure loss when cleaning frequency was reduced. The early claims that baghouses are insensitive to coal/ash properties, inlet grain loading and particle size distribution and that the cleaning cycle can be adjusted to reduce pressure loss are not always valid.

There are some reports that an increase in RA flow rate will sometimes increase pressure loss, again contrary to logic. Increasing RA flow rate will also increase the amount of fines going back to the inlet duct. The foregoing explanation for cleaning frequency may also explain the anomaly for reverse air flow rate.

Reference (4) recommends against increasing the RA flow rate, "as a practical method of reducing pressure drop." The same reference includes a representation, Figure 4, showing the relation between tubesheet pressure loss and RA air-to-cloth ratio. Data for Figure 4 were obtained at the EPRI test facility at Arapahoe. The conclusion of Reference (4), based on these data, that "...pressure drop is not a strong function of reverse-gas flow rate" may be specific for the Arapahoe test baghouse (FFPP) and not necessarily correct for reverse air baghouses in service. Unlike baghouses in service, the FFPP operates at constant flow rate through each of the four compartments when on-line and the RA flow from a compartment being cleaned goes to the plant chimney instead of to the on-line compartments. Figure 4 and this conclusion of Reference (4) may not be valid for RA baghouses in service.

Reference (4) compares RA cleaning with shake/deflate (S/D) cleaning and concludes that, "Reverse-gas cleaning is less effective in removing residual dust cakes and reducing pressure drop than reverse-gas/sonic or shake/deflate cleaning." The comparison favors S/D cleaning. No reasons are given to explain why one cleaning system seems to work better than another. Both RA and S/D use bag deformation and reverse flow (backwash) as cleaning mechanisms. We can postulate two reasons why S/D works better than RA and both are related to the author's premise.

First, we know that RA flow by-passes the upper part of the bags and takes the path of least resistance through the bottom of the bags. It has very little effect on the tops of the bags. S/D on the other hand, applies most of its cleaning effort at the tops of the bags. Assume that the premise is correct and that fines do accumulate in the tops of the bags. The violent agitation at the tops of the bags in S/D cleaning can be expected to cause some of the fine particles to agglomerate and become heavy enough to overcome the stack effect and to fall to the hopper. In other words, S/D cleans better at the top and RA cleans better at the bottom. If the cake in the bag were of uniform composition and distribution, the performance of the two cleaning methods should be more nearly equal. The fact that they are not suggests that the cake is not uniform in composition and/or distribution. If the accumulation of fines in the tops of the bags is a significant cause of pressure loss and if S/D cleaning eliminates more of those fine particles this line of reasoning would seem to substantiate the premise.

Second, the deflation (RA) flow in S/D cleaning is a low flow rate of short duration. Flow from the hoppers back to the inlet duct is thus much smaller than flow from an RA hopper. Thus S/D cleaning recirculates a much smaller amount of fines than RA cleaning. There is, therefore, significantly less loss of filter area due to blinding in the downstream compartments. The net result is that S/D baghouses can operate at a higher cloth ratio for the same pressure loss as compared with RA baghouses. The premise and this reasoning provide an explanation for the difference in performance of these two cleaning methods.

THE INDEPENDENT REVERSE AIR CLEANING SYSTEM

Having used the premise to explain some anomalies of RA baghouse performance there remains the task to overcome the unsatisfactory relations among RA cleaning, cleaning frequency, RA flow rate and the non-uniform performance of the compartments.

These shortcomings of conventional RA cleaning can be avoided by a method that is applicable for both new and existing baghouses. It is called the Independent Reverse Air Cleaning System (IRA). This system was described and recommended in Reference (6). Diagrams of a conventional RA system and an IRA system are shown on Figures 2 and 3.

The IRA system eliminates the recirculation of fines in the baghouse. It should also eliminate blinding in the downstream compartments, reduce tubesheet pressure loss, reduce residual cake pressure loss and equalize compartment collection. The residual cake pressure loss consists of several parts. It includes the entrance loss at the bottom of the thimble and the loss through the cloth, both of which are negligible. The cake loss itself consists of two parts: the cake that adheres to the bag and is not dislodged by RA cleaning; and the floaters which return to the bag as the compartment is returned to service after cleaning. If the customary null period just prior to reinflation is omitted and if the RA flow is continued up to the start of reinflation it will remove a lot of the floaters. With an IRA system this removal of the floaters will reduce residual cake loss.

After a baghouse with a conventional RA system has been in service for some time there will be a thin coating of particulate on the outside of the bags. This may be due to penetration but, more likely, it is probably due to recirculation of fine particles in the "clean gas" RA flow. If the IRA system is arranged to use heated air instead of "clean" gas it will then eliminate the dirtying of the clean side and also whatever additional residual cake pressure loss may be due to the cake on the outside of the bags.

Conventional RA baghouses include the RA flow in the design cloth ratio. This assumes that the RA flow will be equally divided among the on-line compartments and that it is acceptable to provide the same amount of cloth in each compartment. In the foregoing discussion we have seen that compartments 1 and 2 do not get any RA flow and that first No.10 and then No.9 may get all of the RA flow when they are cleaned. Because we have cloth in Nos.1 and 2 that does not have to filter RA flow, these compartments will operate at higher-than-design cloth ratio. They filter at a higher flow rate and collect more particulate than the downstream compartments.

It is not likely that we can find a way to distribute the extra (RA) cloth among the compartments to make all compartments collect the same amount. With the IRA system that will not be necessary. For a new RA baghouse, put enough cloth in the main baghouse to satisfy the design cloth ratio without the RA and with the same amount of cloth in each compartment. Then design a pulse-jet (PJ) baghouse to filter the RA flow. Because the main baghouse may be designed for a cloth ratio of 2 fpm and the PJ baghouse for 4 fpm the total cloth area will be less for the IRA as compared with a conventional RA system.

Application data for outside collectors (pulse jets) are excellent and can be relied upon to provide a high performance, reliable dust collector for the RA flow in an IRA system.

For both new and retrofit installations the IRA will provide several other benefits. It will make it possible to reduce pressure loss by increasing cleaning frequency if necessary. It will reduce pressure loss because eliminating the recirculation of RA particulate is equivalent to reducing inlet grain loading and that, in turn, should make it possible to reduce cleaning frequency. Least frequent cleaning is desirable because it increases the time the baghouse operates at its lowest cloth ratio and therefore at its lowest operating pressure loss. Least cleaning frequency is also desirable because it increases bag life and reduces emissions.

The material collected in the PJ will be different from that collected in the main baghouse. The difference will be in size distribution, bulk density and possibly in chemical properties. The separation may enhance the value of one or both of the two kinds of collection or, it may reduce the cost of waste disposal.

For example, the material collected by a bottom inlet baghouse with a conventional RA system serving a spray dryer FGD system will be a mixture of fly ash, the products of chemical reactions and some unreacted reagent. With an IRA system the material collected by the PJ will be mostly products of completed chemical reactions with very little, if any, unreacted reagent. If the spray dryer/baghouse system includes recycle of the baghouse catch the recycle material will be improved by the removal of the fully reacted material in the PJ.

For a retrofit installation the IRA will also reduce the cloth ratio, and therefore pressure loss, of the main baghouse. That reduction, plus the reduction in inlet grain loading, will reduce cleaning frequency and provide those consequent benefits.

For a new installation, the costs for a conventional RA system and for an IRA system should be nearly equal. The costs for a retrofit IRA would be high and site specific. In some cases, the high costs for a retrofit IRA may be justified by the evaluated cost of reduced pressure loss, assuming that there will be a way to predict the amount of pressure loss reduction. One way to determine the pressure loss reduction at an existing installation would be to install new reverse air dampers on the hoppers, and to arrange ducts and the RA fan so that the reverse air can be discharged into an existing, adjacent, dust collector, or discharged to the plant chimney long enough to run a test. The test should include "before" and "after" data for both pressure loss and hopper collection. The pressure loss reduction determined by the test may have an evaluated cost sufficient to justify installing a complete IRA system.

EXAMPLES OF PRIOR IRA EXPERIENCE

In conventional bottom inlet pressure type RA baghouses the RA fan suction is from the hopper. A small cyclone or baghouse on the suction side of the fan is used to protect the fan and to clean the RA flow before it is returned to the inlet duct. The clean exhaust is usually discharged directly to atmosphere from each compartment without dampering. The RA flow is usually atmospheric air drawn in through the compartment clean exhaust duct.

In conventional suction type RA baghouses the RA flow goes directly from the hopper to the inlet duct and, because it can only go downstream, it will not be recirculated to upstream compartments. In conventional pressurized RA baghouses the RA flow is returned to the head end of the inlet duct and is thus distributed to all of the compartments.

The Alamitos baghouse, first on a utility boiler, is a positive pressure baghouse with RA cleaning. The RA flow is used to clean the bags in a compartment and to empty the hopper at the same time. The flow is cleaned by mechanical cyclones before return to the head ends of the horseshoe-shaped inlet duct.

These examples show that the IRA system is old art for pressure type RA baghouses. Its use for suction type RA baghouses has the potential for significant improvements in baghouse performance.

It is the author's opinion that the pressure loss contribution of small particles has been underestimated and that we have made that contribution more significant by recirculating the fines as we do in conventional reverse air cleaning systems. The independent reverse air cleaning system may be the way to minimize the pressure loss due to the small particles.

RECOMMENDATIONS FOR SOME NEW BAGHOUSE TEST PROCEDURES

There are a lot of things we do not know about fabric filtration and baghouse performance. One reason for that is because our customary testing procedures have some deficiencies and limitations.

Procedures requiring Pitot tube and other traverses are time consuming, labor intensive and not very accurate. It would be very helpful to have certain data, such as flow rate, grain loading and particle size distribution, continuously and simultaneously in several places in a baghouse in service. These data cannot be obtained by traverses.

Conventional intermittent testing has not provided the means to reliably and accurately predict pressure loss. We do not know how grain loading entering the bags varies with time on line. We do not know how much particulate is recirculated in a conventional RA system and what effect that has on pressure loss. We are still debating whether or not fine particles accumulate in the top of bottom inlet bags and also, the relative merits of top and bottom inlet designs. We know very little about filter cake morphology, bulk density, the relation between particle size and porosity, nor the causes and possible cures for cake collapse.

There is a good possibility that some different test procedures may provide the data we need to resolve some of these unknowns. For example, it would be helpful to have certain data versus time on line; for compartments at the head end of the inlet duct versus those at the tail end of the inlet duct and for compartments at the start of the cleaning cycle versus those at the end of it.

The data to be acquired for those combinations should include grain loading entering the compartment and entering the bags. This would tell us how hopper fallout varies for those combinations. It should include isokinetic sampling at bag inlets, tubesheet pressure differential and compartment flow rate. With these data we could determine k_2 and how it varies for those combinations. It would also help us to determine the significance of RA recirculation both for RA flow distribution and RA particulate.

There are some test procedures that have been considered, and to the author's knowledge, not used. These could resolve the question of whether or not fines accumulate in the top of bottom inlet bags; the bulk density of fly ash filter cake in service and, the variation of areal density and particle size versus altitude in the bag.

Some of those tests can be done using a modern type of hot wire anemometer to measure mass flow rates. These relatively new instruments are being used in various industrial applications and by the boiler companies for combustion control. They are very accurate, rugged and can be used in high temperature dirty gas flows. The feature that makes them especially suitable for our testing is that they can be mounted in a duct in an array that provides the required multi-point data simultaneously and continuously instead of intermittently, point-by-point as is the case with traverses. The signals from each point can be processed by a black box to provide an accurate recording of mass flow rate versus time.

Another relatively new instrument is an isokinetic sampling probe which uses two of the new hot wire anemometers. One measures the velocity in the duct and the other measures the velocity in the sampling probe. These two signals are processed by a black box to control the vacuum pump that pulls the sample so that the two velocities are always the same. This instrumentation avoids the inherent inaccuracy on non-isokinetic sampling which is unavoidable when duct and probe velocities are measured intermittently, separately and at different times. This instrumentation makes it possible to sample at a bag inlet to obtain a continuous record of bag inlet velocity and samples of particulate, at intervals, to provide data on grain loading and particle size distribution versus time on line. A suggested arrangement of this probe for bag inlet testing is shown on Figure 4.

In addition to the foregoing test procedures there are two other kinds of testing that require some special apparatus. The first is a "sock test" which requires special bags. This test can provide fly ash samples from different altitudes in the bag. These samples can tell us the rate of collection, the bulk density and the particle size distribution collected at a number of locations from top to bottom of bags in service. The bag design and the test set-up are shown on Figure 5. The bag design has been discussed with one supplier to confirm that these bags are commercially available.

Another test set-up is illustrated by Figure 6. This procedure can be used to determine the face velocity distribution over the length of a bottom inlet bag in service. One purpose of this test would be to see if fines accumulate in the top of bottom inlet bags to an extent that reduces or eliminates flow through the top of the bag. Loss of significant filter area in compartments at the tail end of the inlet duct would confirm the rationale for the independent reverse air cleaning system. A similar test for top inlet bags would help to resolve the relative merits of top versus bottom inlet bags.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

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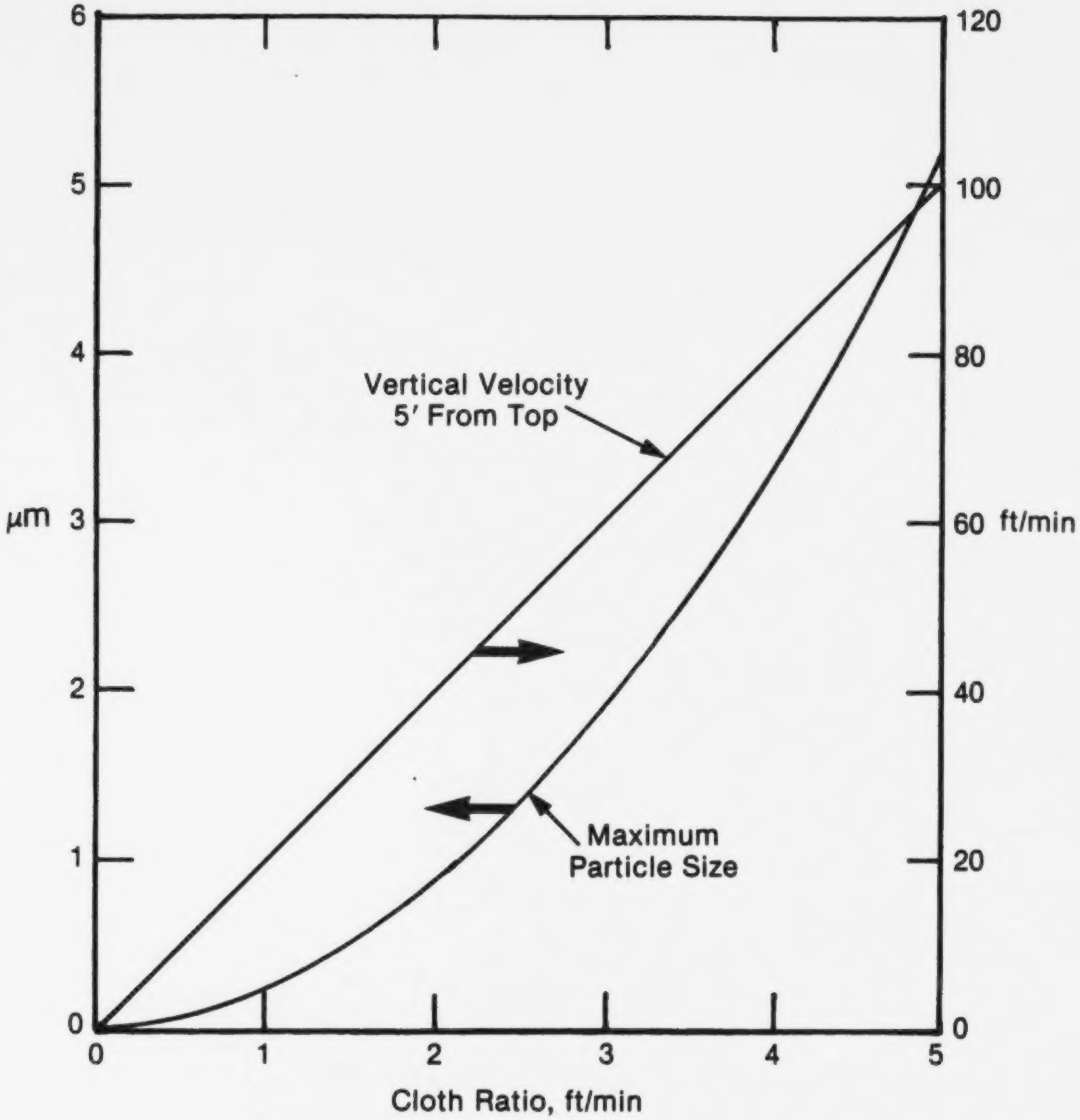

Figure 1. Particle Sizes in Top of Bag

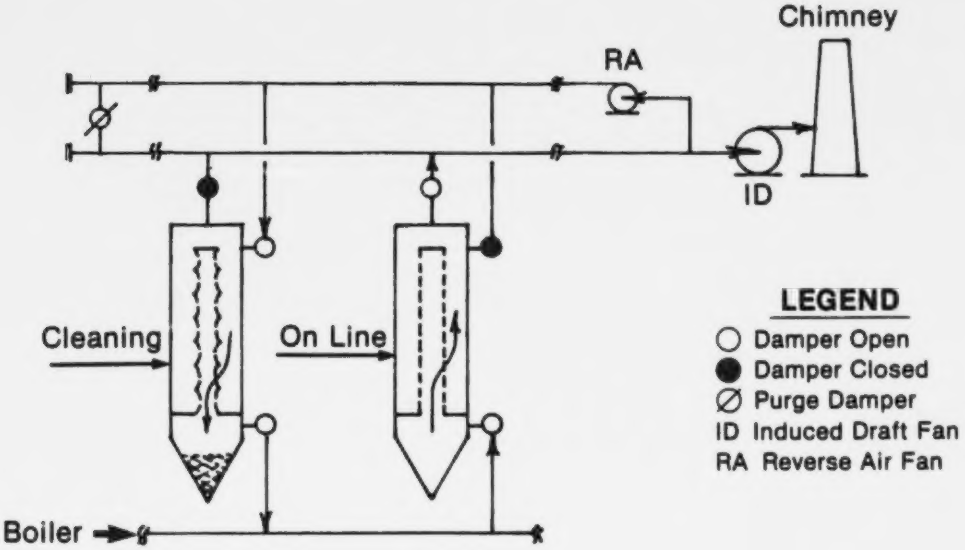

Figure 2. Conventional Reverse Air Cleaning System

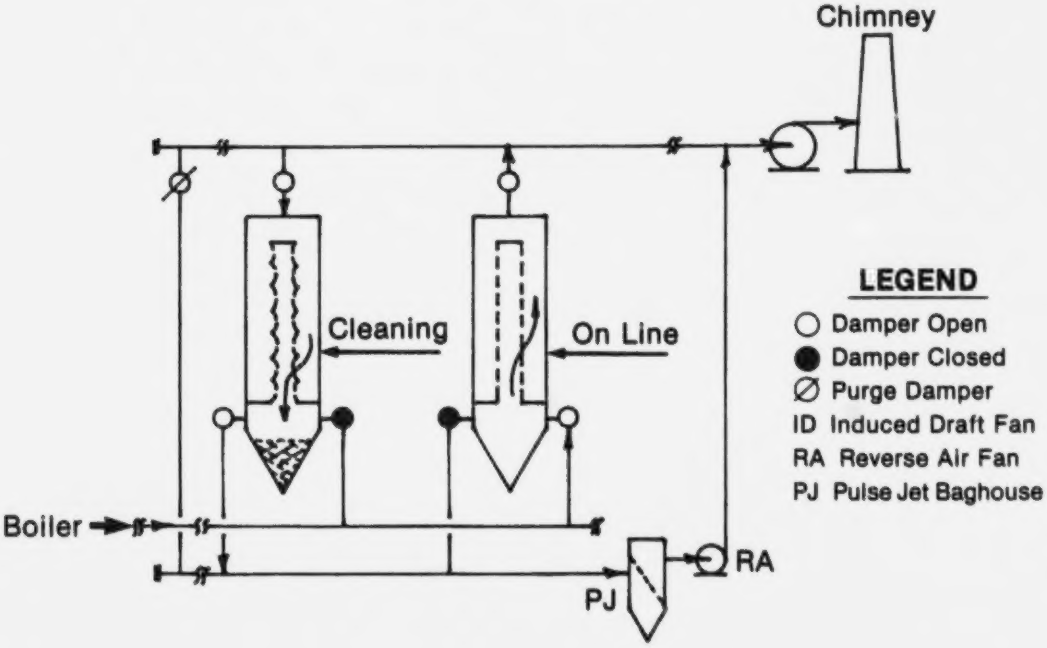

Figure 3. Independent Reverse Air Cleaning System

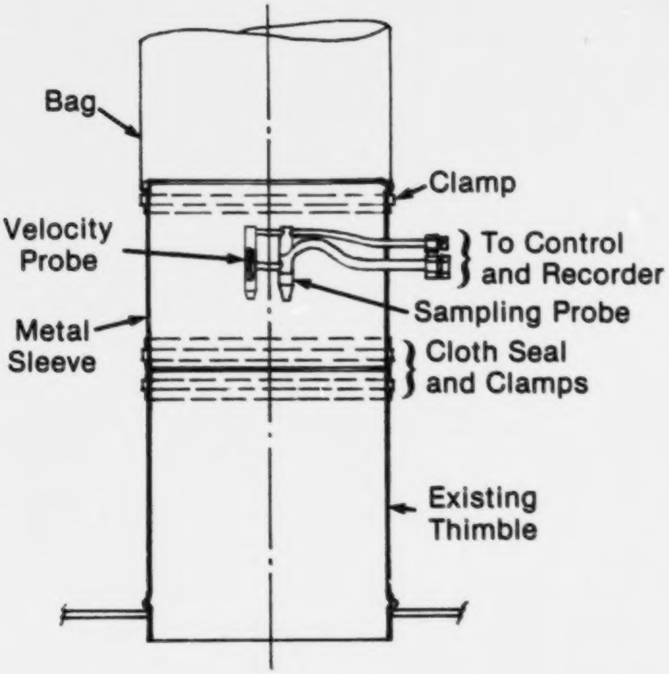

Figure 4. Bag Inlet Test

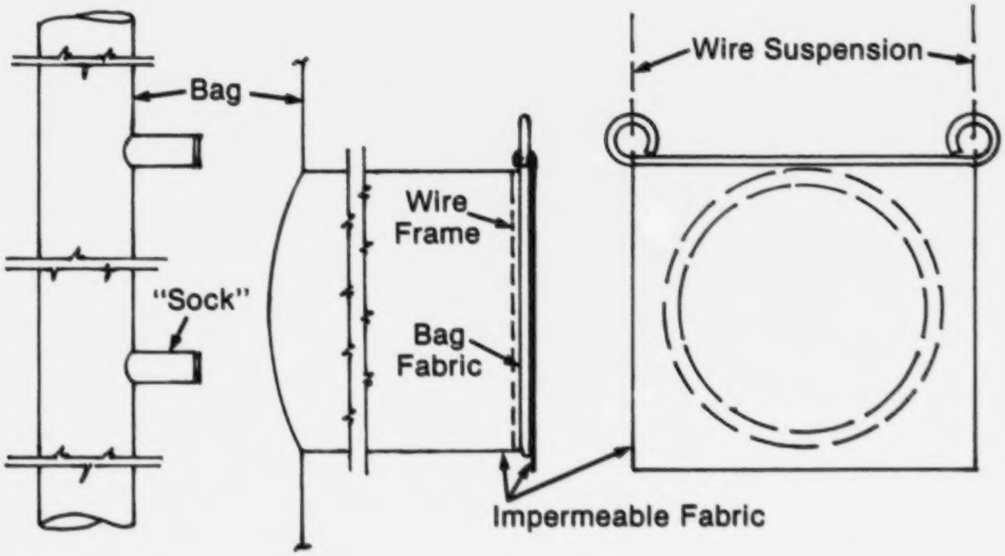

Figure 5. "Sock Test"

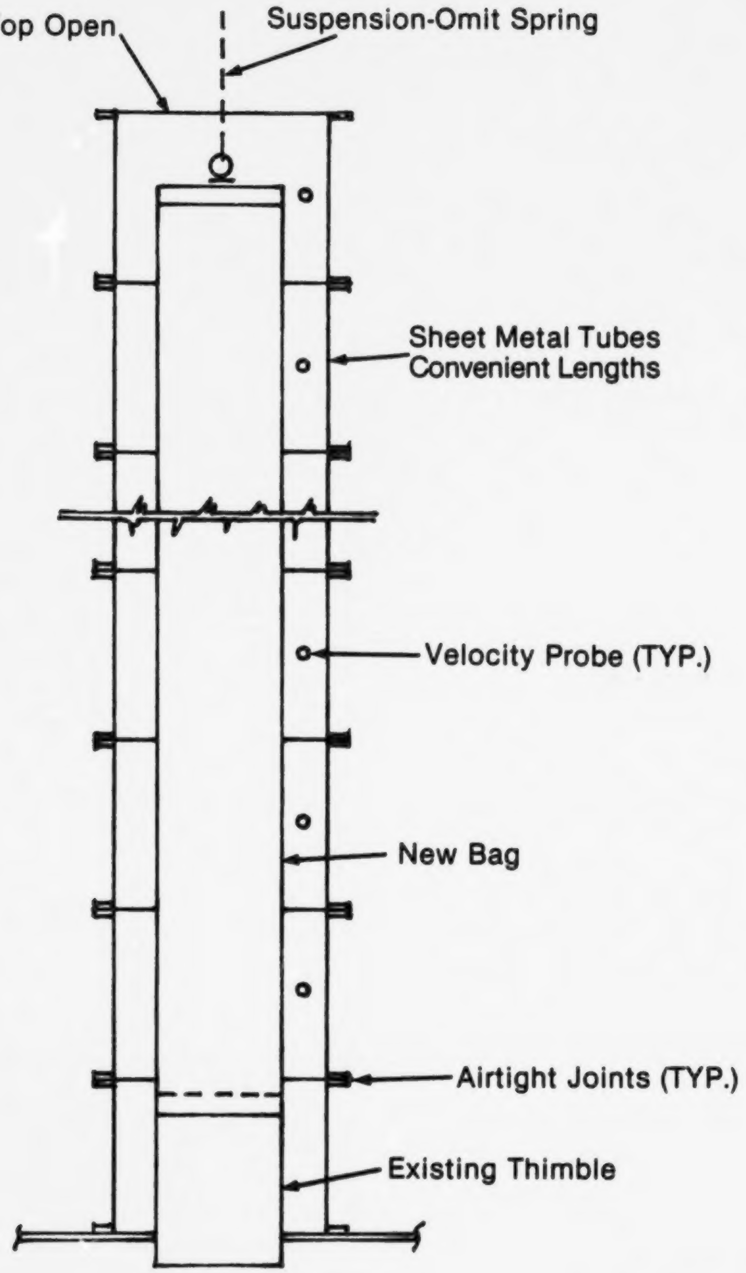

Figure 6. Face Velocity vs Bag Length

TABLE 1. DISTRIBUTION OF REVERSE AIR FINES, POUNDS/CYCLE

Compartment Number										
1	2	3	4	5	6	7	8	9	10	Sum
clean	0	4.5	4.5	4.5	4.5	4.5	4.5	4.5	4.5	36
0	clean	4.5	4.5	4.5	4.5	4.5	4.5	4.5	4.5	36
0	0	clean	0	6	6	6	6	6	6	36
0	0	0	clean	6	6	6	6	6	6	36
0	0	0	0	clean	0	9	9	9	9	36
0	0	0	0	0	clean	9	9	9	9	36
0	0	0	0	0	0	clean	0	18	18	36
0	0	0	0	0	0	0	clean	18	18	36
0	0	0	0	0	0	0	0	clean	36	36
0	0	0	0	0	0	0	0	36	clean	36
0	0	9	9	21	21	39	39	111	111	360

AUSTRALIAN EXPERIENCE WITH FABRIC FILTERS ON POWER BOILERS
AN UPDATE FOR 1987

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ABSTRACT

Previous papers presented to the first, second and third EPRI conference on Fabric Filter Technology for Coal Fired Boiler Plant highlighted the history of the use of fabric filters in the power industry in Australia especially within the Electricity Commission of New South Wales (ECNSW).

Major installations currently utilising shaker cleaned fabric filters for flue gas cleaning within the ECNSW system consist of 4 x 660 MW units at Eraring Power Station, 4 x 660 MW units at Bayswater Power Station and 2 x 660 MW units under construction for Mount Piper Power Station.

Contracts for pulse jet cleaned fabric filters have also been let for 2 x 350MW boilers at Munmorah Power Station where the fabric filters will be retrofitted into existing precipitator casings.

As previously reported, the Eraring units have always operated successfully from an emission control viewpoint, but initially differential pressures were higher and bag life shorter than expected. Since then, developments in fabric surface treatment and cleaning system design which were implemented over 3 years ago, have improved differential pressure to meet all performance requirements and bag life is currently approaching 4 years with further improvements expected.

This paper reviews the investigation program involving both the main plant and an 8 cell pilot plant at Eraring which resulted in these improvements in operating pressure drop and bag life and also looks at the current operation of the fabric filters on the 8 x 660 MW coal fired boilers at Eraring and Bayswater Power Stations.

Finally, the paper gives details of the 2 x 350 MW pulse jet fabric filter retrofit program currently undertaken at Munmorah Power Station.

INTRODUCTION

Several papers presented to the first, second and third EPRI conferences on Fabric Filter Technology for Coal Fired Boiler Plant (1, 2, 3 and 4) have described in detail the design and early operation of the Howden shaker cleaned fabric filters for the 4 x 660 MW coal fired boilers at Eraring Power Station.

The Eraring fabric filters are each designed to handle up to 990 m^3/s (2,097,400 cfm) of flue gas leaving the air preheater with a design temperature of 110°C (230°F). In the event of a thermal excursion, a maximum spot temperature of 150°C (302°F) can be controlled to a maximum spot temperature of 130°C (266°F) by means of air infiltration to allow the use of acrylic filter bags.

The filter for each boiler consists of 40 individually isolatable cells designed to give an air to cloth ratio of 0.01 m/s (2 $acfm/ft^2$) with 10 cells out of service and 1 cell off-line for cleaning.

Performance requirements for these filters were an overall plant flange to flange pressure drop of 2 kPa (8 ins wg.), emission to be less than 50mg/Nm^3 (0.022 gr/Nft^3) and a minimum bag life of 8760 hours.

As previously reported, the Eraring filters have always operated successfully from an emission control view point but, initially, differential pressures were high and although bag life more than doubled the required value, bag life was shorter than expected.

This paper summarises the investigation program involving both the main plant and an 8 cell pilot plant that resulted in fabric surface treatments and cleaning system design changes which, when implemented at Eraring over 3 years ago, resulted in very significant improvements in differential pressure and bag life.

Results of long term pilot scale and full scale comparative tests are discussed.

The fabric filters for the 4 x 660MW coal fired boilers at Bayswater Power Station[1] which are of the same design as Eraring have now operated for between 7,000 and 18,000 hours after commissioning between December, 1987 and December, 1986. The Bayswater Unit 4 fabric filter was the first unit to incorporate all of the design improvements prior to start-up and the performance of this unit is discussed in some detail.

Current performance of the Eraring and Bayswater filters is compared to U.S. utility experience while Australian experience with oil only firing on fabric filters is also discussed.

The final part of the paper concerns the high ratio pulse jet fabric filter retrofit program being undertaken by the ECNSW on 2 x 350 MW coal fired boilers at Munmorah Power Station.

PILOT PLANT INVESTIGATIONS

The pilot plant used for the investigations has been described in an earlier paper (4) and consists of 8 cells with 20 full size bags in each cell. Flue gas is drawn isokinetically from the A-side inlet duct of the Eraring Unit 3 fabric filter.

The pilot plant has operated at Eraring for 5 years and accumulated 25,000 hours of operation.

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[1] Gas volumes and cloth areas at Bayswater are approximately 5% higher than for Eraring.

Investigations have focused on the following areas:

Filter Bags - over 100 filter bag designs, with variations in surface finish, permeability, fabric structure etc. were tested.

Cleaning Mechanism Hardware - several new methods of attaching the filter bags to the shaker racks were evaluated.

Shaker Cleaning Parameters - investigations were carried out in a specially equipped cell to evaluate the effect of varying shaker amplitude and frequency over a wide range of values.

Sonic Augmentation - a cell was equipped with a 250Hz nominal frequency sonic horn to determine any improvement in cleaning effectiveness.

As a result of the pilot plant investigations a design capable of meeting all performance requirements was developed in 1984. The long term performance of this design which combines a fabric surface treatment with a rigid cap-top bag connection is compared to the pilot plant performance of the original design in Figure 1. At an air to cloth ratio of 0.01 m/s (2 acfm/ft^2) average cell-plate pressure drop was reduced by over 75% to 0.6 kPa (2.4 ins wg.) which was approximately 60% of the original design value. Emission did not significantly increase.

During the pilot plant investigations several new fabric designs capable of significant further improvements in performance were developed. The performance of some of these new fabric designs is shown in Figures 2 and 3. As well as improving the pressure drop performance the new designs were shown to significantly lower the emission levels such that emission levels of 5 mg/Nm^3 or less for full scale filters could be expected.

FULL SCALE INVESTIGATIONS

In parallel with pilot plant investigations an extensive series of investigations were conducted on the Eraring filters.

FLY ASH PHYSICAL AND CHEMICAL PROPERTIES

Numerous physical and chemical analyses of the fly ash from the Eraring fabric filters and the similarly designed Tallawarra filters[1] were carried out to determine the relationship between fly ash characteristics and fabric filter performance.(4)

Factors which were found to correlate with higher pressure drop or drag were increased shear strength of the fly ash layer, larger number of spheroidal rather than angular fly ash particles and increased proportion of Ca, Mg and Na in the fly ash.

CLEANING SYSTEM OPERATING PARAMETER INVESTIGATIONS

An extensive program to determine the effect of variations to the basic cleaning system operating parameters on average cell-plate pressure drop was carried out on several of the Eraring fabric filters prior to retrofitting the new design improvements.

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[1] Although of a similar design and collecting fly ash from the combustion of similar coals to Eraring, the Tallawarra filters were operating with significantly lower pressure drop.

Variations were made to pre-shake dwell time, shaker on-time, post-shake dwell time, shaker speed and amplitude and filtering cycle on-time.

Although the investigations were carried out in a controlled manner the variations in the day-to-day operation of the operating boiler made the results of these investigations inconclusive except where a marked increase in shaker speed, amplitude or on-time improved performance.

BAG WEIGHT INVESTIGATIONS

From the start of operation of the first unit at Eraring regular bag weight checks were carried out.

The rapid rise in pressure drop on Unit 1 was mirrored in the rapid rise of bag weight in all cells.

However, when Eraring Unit 4 fabric filter was commissioned an opportunity arose to determine if improved pressure drop or drag was coincident with a lowering of bag weights. Surface treated tab-top bags were installed in 29 cells, standard tab-top bags in 10 cells and surface treated cap-top bags in 1 cell. These results are compared as areal dust cake loadings in Figure 4. The relative areal dust cake loadings reflect the differences in pressure drop between the different designs shown in Figure 1 indicating that the high pressure drop at Eraring was related to the inability of the original filter bag/cleaning system design to adequately remove the dust cake from the filter bags.

COMPARISON OF PILOT SCALE AND FULL SCALE PERFORMANCE

The use of pilot scale fabric filters to predict the performance of full scale plant is well established.

At Eraring it was possible to operate some cells of the 8 cell pilot plant in parallel with the Unit 3 filter over a long period with the same filter bags, cleaning mechanism and cleaning system operating parameters.

The pressure drop and emission for the surface treated cap-top acrylic bags are compared for pilot and full-scale in Figures 5 and 6 respectively.

From Figure 5 it can be seen that pilot scale tests can reliably predict full scale plant pressure drop or drag. This result is supported by tests carried out at Eraring by Howden on other filter bag designs at pilot and full scale.

From Figure 6 it can be seen that emission tests from the pilot plant were only indicative of emission from the full scale plant. This result is supported by tests carried out at Eraring on other filter bag designs at pilot and full scale where pilot scale emission results tended to consistently underestimate full scale results by 40-50%.

CURRENT PERFORMANCE OF BAYSWATER UNIT 4 FABRIC FILTER

The final filter bag and shaker mechanism design developed for the Eraring Bayswater and Mt. Piper fabric filters has now been installed or has been retrofitted to 8 of the 10, 660 MW units with the last two units to be converted by early 1989.

Bayswater Unit 4 fabric filter was the first filter to enter service with the new design from initial start-up. This filter has operated for over 7,000 hours and the pressure drop trace for this period is shown in Figure 7.[1] Performance of this filter and all other retrofitted filters comply with all performance requirements.

Operation of the new design has been so successful that filtering cycle times have been increased from 1 hour to 5 hours at full load on some filters.

Emission for all filters has consistently been below 20mg/Nm^3 (9gr/Nft^3) corresponding to a stack opacity of less than 5% and a clear stack has always been achieved.

The four Bayswater fabric filters have now operated for between 7,000 and 18,000 hours and out of the 200,000 bags only 45 have failed due to accidental installation damage with no further failures during operation.

Bag life on the retrofitted filters is now approaching 4 years with further improvements expected from the Bayswater filters.

Current performance of the Eraring and Bayswater fabric filters compares favourably with U.S. experience as shown in Figure 8 adapted from Smith and Carr (5). Average cell plate pressure drop of 0.9kPa (3.6 ins wg.) at an air to cloth ratio of 0.01 m/s (2 acfm/ft^2) for Eraring and Bayswater is significantly lower than the values shown for both pilot and full-scale filters in the U.S.

OIL FIRING EXPERIENCE WITH FABRIC FILTERS

No bypass of the fabric filter is allowed under Australian environmental regulations during boiler start-up, shut-down, emergencies or in any other operating condition.

During hot and cold boiler start-ups the flue gas generated from oil-only firing at low loads is passed through the filters for periods of up to 24 hours.

Supplementary oil firing is used regularly at low boiler load.

No deleterious effects of these procedures have been found after detailed continuous monitoring and analysis over a six year period.

MUNMORAH RETROFIT PROGRAM

In addition to the shaker type fabric filter experience discussed in the foregoing, Australian utilities have also gained considerable experience in applying pulse jet fabric filters to upgrade the emission control of existing coal fired boilers.

Because of their ability to operate with a much higher air to cloth ratio of up to 0.0229 m/s (4.5 acfm/ft^2) on fly ash and thus the ability to offer considerable cost and space advantages, pulse jets have been retrofitted into or in place of existing precipitators on a number of power stations. Since 1975, electrostatic precipitators on boilers ranging in size up to 60 MW with a total of 950 MW of generating capacity have been converted to pulse jet fabric filters.

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[1] Longer filtering cycles used at Bayswater mean average cell plate pressure drop is higher than shown earlier for Eraring pilot and full-scale filters.

More recently the ECNSW decided to apply this technology to 2 x 350 MW pulverised coal fired boilers as part of a life extension program at Munmorah power station. Following extensive pilot trials the ECNSW awarded the contract for Unit 4 boiler conversion to Flakt and Unit 3 boiler to James Howden Australia. These will be the largest pulse jet cleaned fly ash filters in the world.

The conversion of Unit 3 precipitator will be based on a design gas flow of 528 am^3/s (1,118,620 acfm) with a maximum temperature of 155°C (310°F). Air infiltration will be used to control the flue gas temperature to 130°C (266°F) to allow the use of acrylic bags. The inlet dust burden is in the range of 25-40g/Nm^3 (11-17 gr/Nft^3) with a median particle size of 5-30 microns. Allowable pressure drop across the fabric filter is 2.5 kPa (10 in.wg) with an air to cloth ratio of 0.0203 m/s (4 acfm/ft^2). The specified emission guarantee is 50 mg/Nm^3 (0.022 gr/Nft^3) although much lower emissions and a clear stack are indicated by the pilot plant results.

The Howden design is based on minimum modification of existing equipment and structures. Essentially, the precipitator casings, hoppers, structures, ducting and insulation will be re-used. An internal division wall, transverse to the gas flow, will be added to the casings to provide 8 compartments per boiler so that maintenance can be carried out in any compartment without impeding boiler output, Figure 9. However, normal bag cleaning will be in the on-line mode.

The upper precipitator casing will be modified to incorporate the tube sheet and clean gas outlet which doubles up as a penthouse for bag inspection and replacement. Each compartment will contain 6 modules with 276 bags per module. Bags will be 6.1m (20 ft.) long and each module is cleaned by a rotating manifold to which cleaning air of 80 kPa (12 psig) is supplied by a Roots type blower, Figure 10. Inlet and outlet precipitator evases will be modified to cater for the new 8 compartment filter configuration, and to incorporate gate type dampers for compartment isolation and protection of the casings against excessive pressure excursions. Aerodynamic model tests were carried out to ensure that flue gas entering the casings and flowing upwards into the bag arrays is distributed evenly to avoid bag abrasion and to minimize re-entrainment during on-line cleaning.

In addition to a 0.6 MW pilot plant at Munmorah, Howden has installed a full scale 10MW module at Liddell power station to confirm final design parameters, in particular bag design. Furthermore, the ECNSW is co-funding a 2-year program with EPRI to conduct studies on 2 other 1 MW Howden pilot plants at Eraring power station which are configured to represent a number of generic pulse jet cleaning systems. These trials are the subject of a separate paper during this conference.

The Munmorah Unit 4 pulse jet fabric filter is scheduled to be commissioned in May, 1988 and Unit 3 in May, 1989.

Experience with pulse jets at Munmorah is expected to pave the way for similar retrofits of even larger boilers and also for consideration of pulse jet technology as a viable option for new boilers.

SUMMARY

Extensive pilot and full-scale studies have shown that changes in the basic design of filter bags including surface treatment, permeability and fabric structure can significantly affect the pressure drop and emission of a fabric filter.

Certain physical and chemical properties such as shear strength of the fly ash layer, particle shape and size and the proportion of Ca, Mg and Na in the fly ash would appear to be indicators of the filtration characteristics of a particular fly ash.

Representative pilot plant trials were found to provide a reliable prediction of full scale fabric filter pressure drop or drag and an indication of expected emission.

With the new design improvements implemented over 3 years ago, Eraring and Bayswater fabric filters meet all contractual performance requirements. Bag life is currently approaching 4 years with further improvements expected.

Current performance of these filters compares favourably with U.S. experience.

These new developments in shaker cleaned fabric filter technology will allow significant increases in air to cloth ratios up to 0.015 m/s (3 acfm/ft^2) for new plant complying with current and more stringent future emission requirements in Australia and the USA.

Experience has shown that oil only firing with fabric filters is possible under certain circumstances for extended periods.

Although shaker cleaned fabric filters continue to be preferred for large, new boilers at this stage, pulse jet fabric filters are making rapid inroads and have already been committed for large (2 x 350 MW) power boiler retrofits.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

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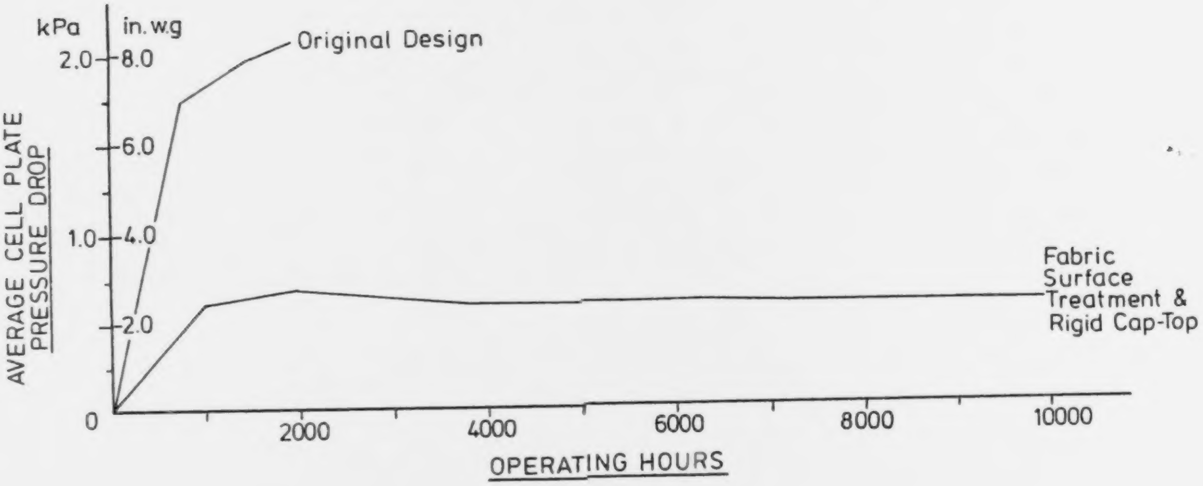

Figure 1. Comparison of Average Cell Plate Pressure Drop (Pilot Plant results) for the Original Eraring Design and the Improved Design with Fabric Surface Treatment and Rigid Cap-Top. Air to cloth ratio is maintained at 0.01 m/s (2 acfm/ft^2).

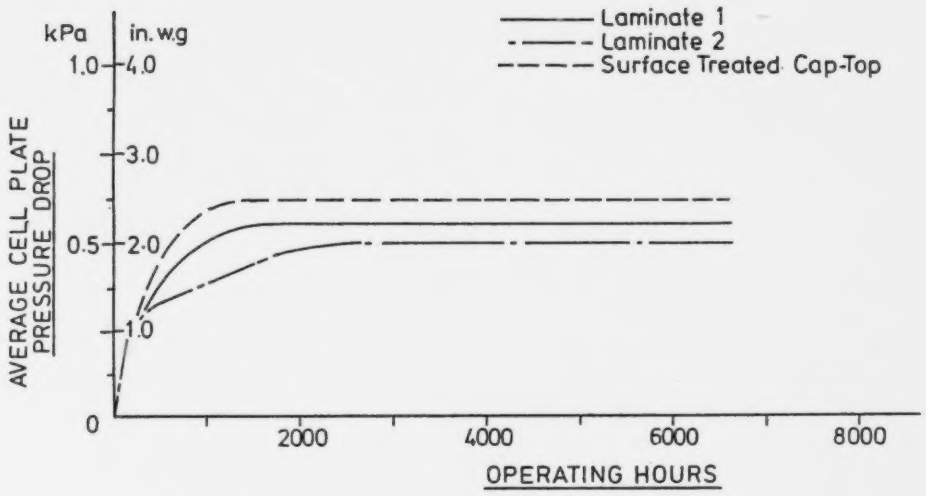

Figure 2. Comparison of Average Cell Plate Pressure Drop for New Fabric Designs developed during Eraring Pilot Plant Test Programme.

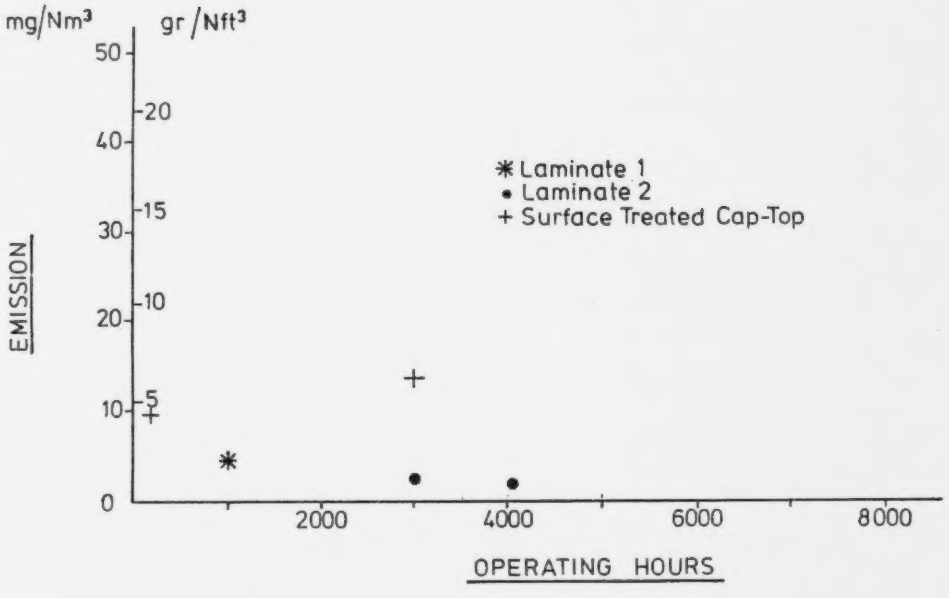

Figure 3. Comparison of Emission for New Fabric Designs developed during Eraring Pilot Plant Test Programme.

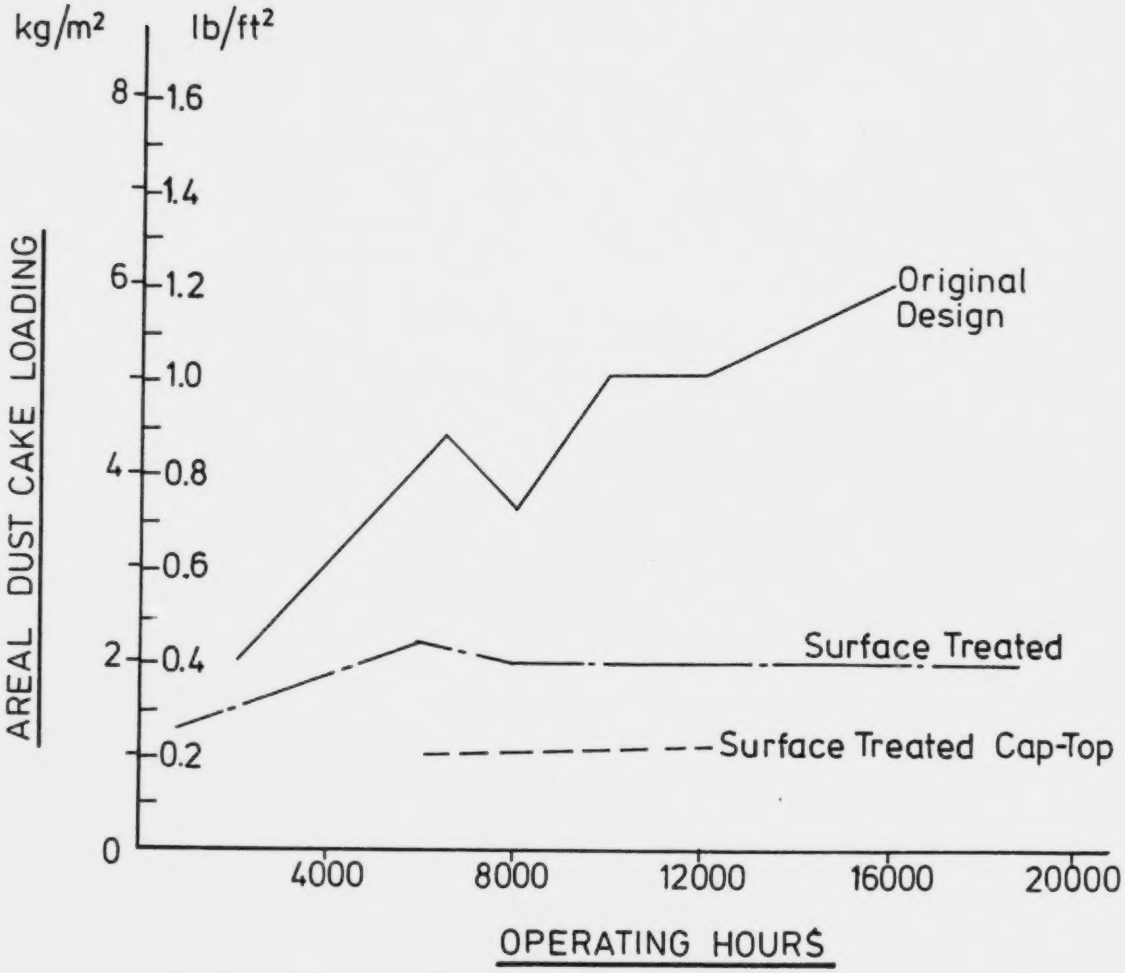

Figure 4. Comparison of Areal Dust Cake Loading for the Original Eraring Design, Surface Treated Bags and the Improved Design with Fabric Surface Treatment and Rigid Cap-Top based on in-situ measurements, Eraring Unit 4 Fabric Filter.

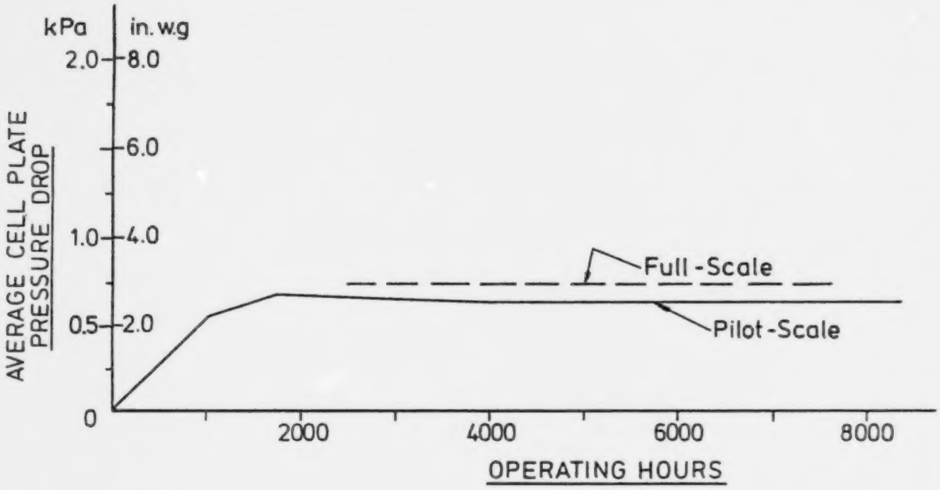

Figure 5. Comparison of Pilot and Full-Scale Average Cell-Plate Pressure Drop for Surface Treated Cap-Top Acrylic Filter Bags.

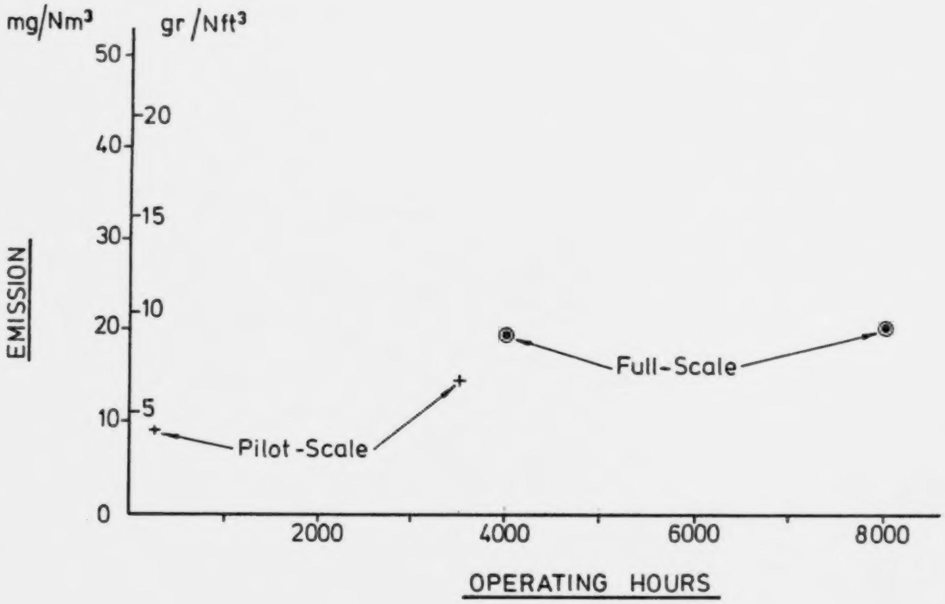

Figure 6. Comparison of Pilot and Full-scale Emission for Surface Treated Cap-Top Acrylic Filter Bags.

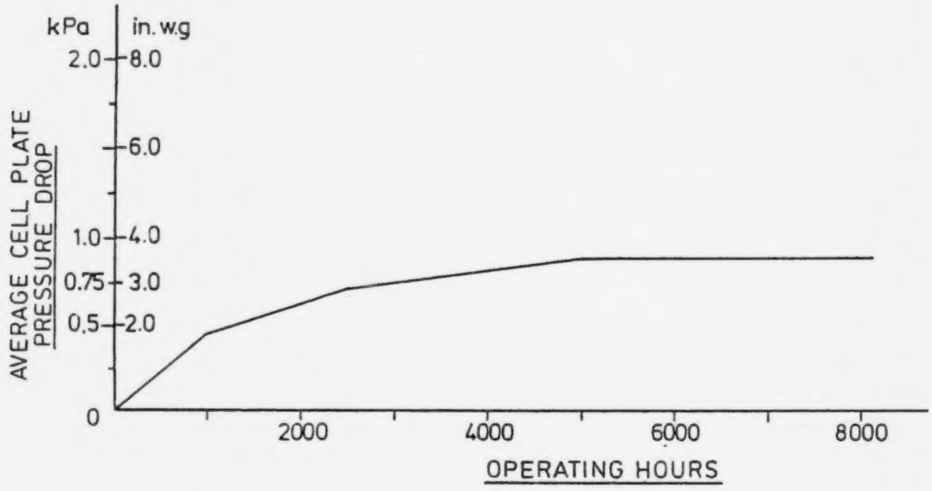

Figure 7. Average Cell Plate Pressure Drop for Bayswater Unit 4 Fabric Filter from Start-up.

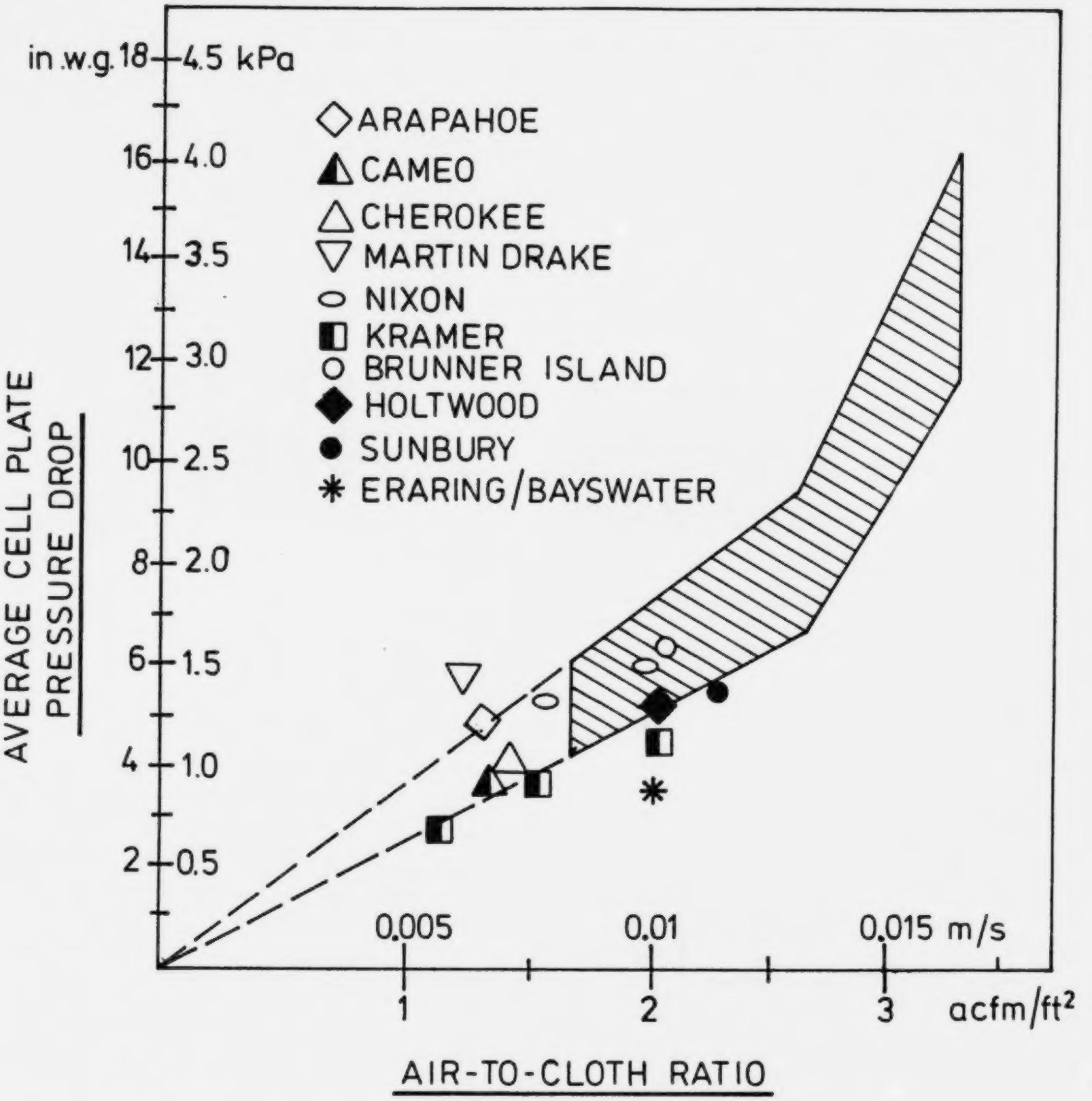

Figure 8. Comparison of the pilot-scale data (shaded area and dashed lines) with data from full-scale reverse-gas baghouses (points) from Smith and Carr (5) and Current Eraring/Bayswater shaker cleaned fabric filter performance.

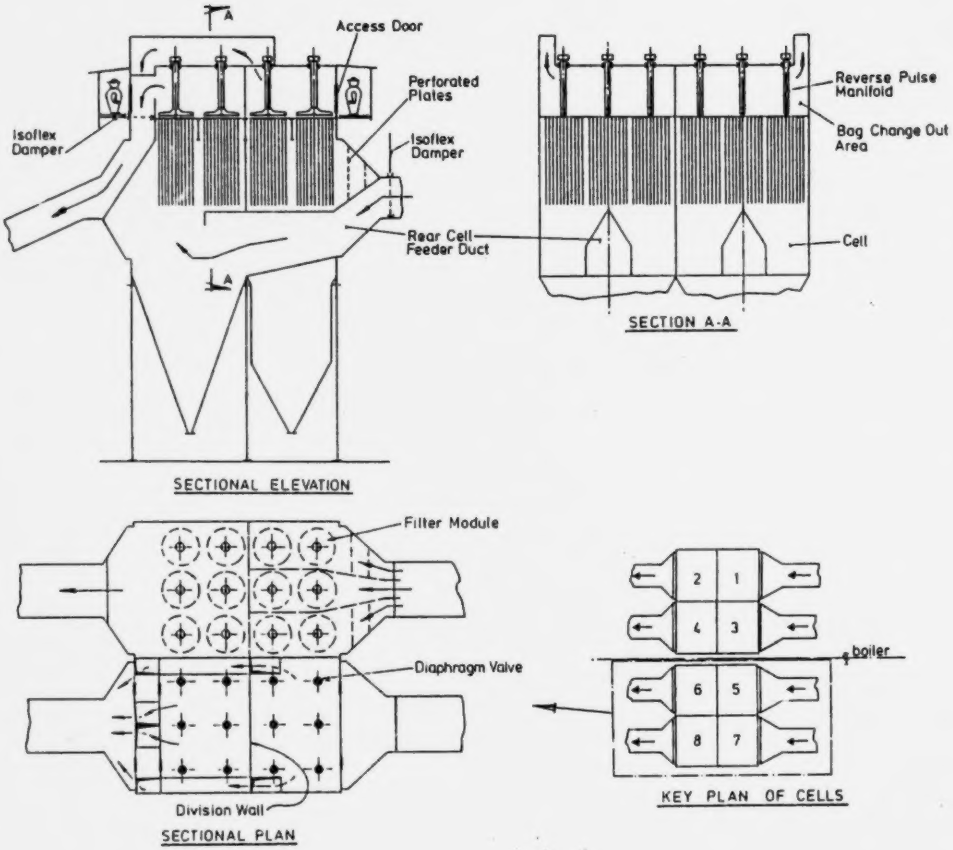

Figure 9. Arrangement of Howden Pulse Jet Fabric Filter Modules Retrofitted into existing Electrostatic Precipitator Casing Munmorah Unit 3.

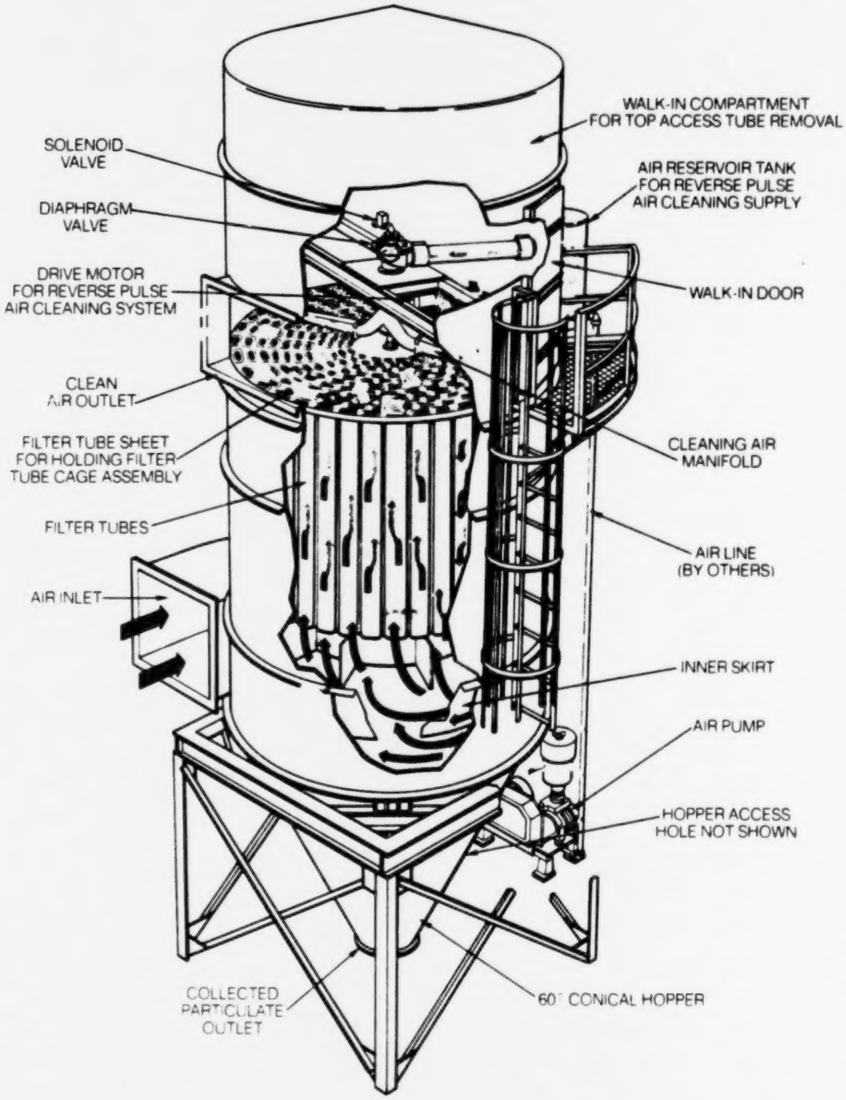
WALK-IN COMPARTMENT FOR TOP ACCESS TUBE REMOVAL
SOLENOID VALVE
AIR RESERVOIR TANK FOR REVERSE PULSE AIR CLEANING SUPPLY
DIAPHRAGM VALVE
DRIVE MOTOR FOR REVERSE PULSE AIR CLEANING SYSTEM
WALK-IN DOOR
CLEAN AIR OUTLET
FILTER TUBE SHEET FOR HOLDING FILTER TUBE CAGE ASSEMBLY
CLEANING AIR MANIFOLD
FILTER TUBES
AIR LINE (BY OTHERS)
AIR INLET
INNER SKIRT
AIR PUMP
HOPPER ACCESS HOLE NOT SHOWN
COLLECTED PARTICULATE OUTLET
60° CONICAL HOPPER

Figure 10. Illustration of a Typical Howden Pulse Jet Fabric Filter Module. For the Munmorah Retrofit, Modules do not have individual casings and hoppers; the precipitator casing provides enclosure.

EPRI HIGH-SULFUR SPRAY DRYER/FABRIC FILTER PILOT RESULTS

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ABSTRACT

In order to evaluate spray drying as a potential high-sulfur flue gas desulfurization (FGD) technology for both new and retrofit installations, EPRI has built and is operating a 4-MW spray dryer/fabric filter pilot unit at their High Sulfur Test Center in Barker, New York. This paper discusses the results of the first nine months of operation of the pilot unit.

The design and operation of the unit is intended to simulate a full-scale utility high-sulfur spray dryer/fabric filter system. The fabric filter has five compartments, each with up to 36 eight inch by 22 ft. woven fiberglass bags. It operates at a filtering air-to-cloth ratio of 1.6 to 1.7 acfm/ft^2, with reverse-gas cleaning.

This paper will emphasize results from initial operation of the fabric filter downstream of the high-sulfur spray dryer. The results presented show that the system is capable of SO_2 removal levels in excess of 90 percent and that particulate emissions remain well below 0.03 lb/10^6 Btu. Fabric filter pressure drop and bag weights remain acceptably low. However, a few adverse effects of the spray dryer on fabric filter operation are discussed.

INTRODUCTION

There are 12 utility spray dryer/fabric filter FGD systems in operation in the U.S., treating flue gas from over 4000 MW of generating capacity. All of these units are installed on boilers that fire low- to medium-sulfur-content coal. A short-term demonstration co-funded by EPRI and the U.S. EPA in 1983 documented that a utility spray dryer/fabric filter FGD system was capable of achieving 90 percent SO_2 removal when treating flue gas from the firing of a high-sulfur-content (3.2 percent sulfur) coal (1). However, there has been no commercial application of spray dryer/fabric filter FGD technology to a utility boiler that fires a high sulfur coal.

Pilot-scale testing of spray dryer/fabric filter FGD technology has been conducted at the EPRI Arapahoe test facility at simulated high-sulfur-coal flue gas conditions, by spiking SO_2 into flue gas from a low-sulfur-coal-fired source. However, there are always concerns that differences between low-sulfur-coal firing and actual high-sulfur-coal firing (such as ash characteristics and flue gas SO_3 content) may influence spiked test results. Furthermore, reverse-gas fabric filters installed in high-sulfur-coal flue gas applications have been noted to develop much heavier residual dust cakes on the bags than those in low-sulfur-coal flue gas applications (2). There is some question as to whether fabric filters in high-sulfur spray dryer applications would have similar tendencies. For these reasons, it is important that high-sulfur spray dryer/fabric filter FGD technology be evaluated with a pilot unit that has a source of flue gas from actual high-sulfur-coal firing.

EPRI has constructed and is operating a facility dedicated to research on FGD systems for high-sulfur-coal-fired utility boilers. The facility, called the High Sulfur Test Center, is located at the New York State Electric and Gas Co. Somerset station in Barker, New York. Other sponsors of the facility include the Empire State Electric Energy Research Corp., New York State Energy Research and Development Authority, Consolidation Coal Co., and the U.S. Department of Energy. The facility includes a 4 MW wet scrubber pilot unit, a 0.4 MW wet scrubber mini-pilot unit, and the subject of this paper, a 4 MW spray dryer FGD pilot unit. The Somerset station boiler provides a source of flue gas for the facility. The facility began start-up in April 1987, and test programs began in June 1987.

There are two pilot-scale particulate control devices currently available at the facility: a five-compartment fabric filter and a five-field ESP. A pulse-jet fabric filter is also currently being installed for future research. The facility is configured so that either the fabric filter or the ESP can treat flue gas from the spray dryer, while the other provides particulate removal for the flue gas fed to the wet scrubbers. To date, all of the operation has been with the fabric filter treating flue gas from the spray dryer and the ESP treating gas that is sent to the wet scrubbers.

The spray dryer/fabric filter pilot unit provides an opportunity to characterize spray dryer and fabric filter performance under actual high-sulfur-coal flue gas conditions. This will prove to be very useful design information for utilities that are contemplating new or retrofit installations of this technology to power plants firing high-sulfur coal.

In the following sections of the paper, a description of the spray dryer/fabric filter FGD process is provided, the pilot unit is described, then results are presented from start-up through February 1988. The results presented document SO_2 removal, particulate control performance, and fabric filter pressure drop and bag weights. Also, fabric filter compartment wall corrosion tendencies and other impacts of the spray dryer on the fabric filter are discussed.

PROCESS DESCRIPTION

Figure 1 is a simplified process flow diagram for the spray dryer/fabric filter pilot unit. The flue gas fed to the pilot unit is isokinetically extracted from the air heater outlet duct on the NYSEG Somerset station boiler. A small portion of air heater inlet flue gas is also extracted as needed for blending to raise the temperature of the slipstream to desired values. The normal sulfur content of the coal fired at the NYSEG station (2.5 to 3 percent) produces a flue gas SO_2 concentration of about 1800 ppmv. The SO_2 concentration of the spray dryer inlet gas is raised to 2000 ppm by spiking with a small amount of SO_2 evaporated from a liquid storage tank. This spiking system also allows tests to be conducted at higher inlet SO_2 levels (2500 to 4000 ppmv). An electric resistance heater is used to make final adjustments to the flue gas temperature. Flue gas enters the spray dryer, where it is contacted with finely atomized slurry that contains fresh lime and recycled solids. The water content of the slurry droplets is evaporated, cooling the flue gas, while simultaneously the lime in the slurry reacts with the flue gas SO_2. Flue gas leaves the spray dryer from the side of the vessel, after entering an outlet duct which is pointed downward and centered in the bottom cone. A portion of the dried solids and normal fly ash drop out in the spray dryer bottom as the flue gas turns to enter this outlet duct, but the majority of the entrained solids are carried out of the spray dryer and are collected in the reverse-gas fabric filter. The entrained solids continue to react with flue gas SO_2, and in fact, SO_2 removal in the fabric filter contributes significantly to the system SO_2 removal. Gas leaving the fabric filter passes through an induced draft fan, then returns to the Somerset station ductwork.

The slurry fed to the atomizer is a mixture of lime, recycle solids, and makeup water. The rate at which this combined slurry is fed to the atomizer is controlled to reduce the flue gas temperature at the spray dryer outlet to within a desired approach to adiabatic saturation. The lime makeup to this feed slurry is produced as a 24 weight percent solids content slurry by slaking pebble quicklime in a paste slaker. The lime slurry content of the atomizer feed slurry is controlled to result in the desired reagent ratio. The reagent ratio is defined as the molar ratio of fresh calcium hydroxide in the atomizer feed slurry to moles of SO_2 in the flue gas entering the spray dryer. Makeup water is added to the feed slurry because at a given reagent ratio the amount of water associated with the lime slurry is generally not enough to reduce the flue gas temperature to the desired value. Solids collected in the fabric filter are recycled to raise the solids level in this combined slurry to 40 weight percent. This recycle is implemented both to improve overall lime sorbent utilization, and to improve droplet drying characteristics. Fabric filter solids are recycled rather than those which drop out in the spray dryer because the fabric filter solids contain fewer oversize chunks that tend to cause pluggage in the atomizer feed system.

PILOT UNIT DESCRIPTION

The pilot unit is sized to treat 12,800 acfm of flue gas at 300°F. The spray dryer vessel has a 13 ft. diameter and a 11-1/4 ft. straight side, with a 45° cone bottom. The vessel has a volume of 2160 ft^3, which results in an 11 to 12 second flue gas residence time when cooling the 12,800 acfm of gas to within a 20°F approach to adiabatic saturation. There is a single rotary atomizer at the top of the vessel. The atomizer has a 50-hp motor, and is equipped with a 250-mm (9.8 in.) diameter atomizer wheel with eight 6-mm (0.24 in.) diameter nozzles. At the normal operating speed of 14,000 rpm, the tip speed at the outer edge of the wheel is about 600 ft./sec.

The fabric filter has five compartments, each with up to 36 bags in a 6 x 6 array. The woven fiberglass bags are 8 inches in diameter and 22 ft. long, are constructed of a 13.5 oz/yd^2, 3 x 1 twill fabric, and have an acid-resistant finish. The fabric filter has provisions for either reverse-gas or shake-deflate cleaning, but all testing to date has employed reverse-gas cleaning. The time between cleanings for each compartment is set at 90 minutes. The cleaning cycles are staggered so that one compartment cleans every 18 minutes, rather than cleaning all five compartments in succession after 90 minutes. The cleaning cycle for each compartment consists of an initial 45-second null period, 30 seconds of reverse-gas flow, and a final 30-second null period. The reverse-gas flow rate is controlled to result in a reverse-gas air-to-cloth ratio of 1.5 acfm/ft^2.

With all 36 bags in each of the five compartments and a flue gas flow rate which results in the design 12-second residence time in the spray dryer vessel, the air-to-cloth ratio at the fabric filter is about 1.4 acfm/ft^2. In August 1987, three bags were removed and the corresponding tubesheet thimbles were capped in each compartment in order to increase the air-to-cloth ratio to about 1.6 acfm/ft^2. This value was thought to be more typical of full-scale utility practice than the initial lower value.

Measurements made on the fabric filter include flue gas flow rate, flange-to-flange pressure drop, individual compartment tubesheet pressure drop, inlet and outlet flue gas temperatures, reverse-gas flow rate, inlet and outlet flue gas SO_2 and O_2 concentrations, and outlet opacity. After several months of operation, orifices were added at the outlet of each compartment in order to allow direct measurement of the flue gas flow rate through individual compartments, and thermocouples were added to allow measurement of compartment wall temperatures.

RESULTS

A number of results from the pilot unit test program are presented. First, typical SO_2 removal performance results across the entire spray dryer/fabric filter system and specifically across the fabric filter are presented and discussed. Then, data which document the performance of the fabric filter as a particulate control device are presented. Finally, a discussion is provided on the observed impacts of the spray dryer upstream on the operation of the fabric filter.

SO_2 REMOVAL

Figure 2 summarizes typical SO_2 removal performance across the spray dryer/fabric filter system. The data are presented as SO_2 removal as a function of reagent ratio. The reagent ratio is defined as the moles of calcium in the fresh lime in the atomizer feed slurry (ignoring any excess lime content in the recycle solids) divided by the moles of SO_2 in the spray dryer inlet flue gas. These data are for slaked Mississippi Lime reagent and reflect operation of the spray dryer at a 300°F inlet temperature and a 2500 ppmv inlet SO_2 content, and for adding sufficient slurry to result in a 20°F approach to adiabatic saturation. Three curves are illustrated. The top curve represents overall system SO_2 removal performance, and the curve just below it represents the percent removal across the spray dryer vessel. The bottom curve represents the contribution of the fabric filter to overall SO_2 removal. This is calculated by dividing the decrease in flue gas SO_2 content across the fabric filter by the system inlet SO_2 content. The fabric filter SO_2 removal is expressed in this manner in Figure 2 so that the spray dryer percent removal and the fabric filter contribution may be summed to get the overall system SO_2 removal.

The results in Figure 2 illustrate SO_2 removal performance for reagent ratio values from about 1.0 to 1.6. The measured system SO_2 removal varies from about

70 percent at a 1.0 reagent ratio to about 93 percent at a 1.6 reagent ratio. A removal level of 90 percent is achieved at a reagent ratio of 1.4 to 1.5.

The dashed line included in Figure 2 represents 100 percent lime utilization. That is, the dashed line represents the SO_2 removal that would result if all of the lime added reacted to remove flue gas SO_2. At a reagent ratio of 1.0, 100 percent lime utilization would correspond to 100 percent overall SO_2 removal. The lime utilization values measured vary from about 70 percent at a reagent ratio of 1.0 to 58 percent at a reagent ratio of 1.6.

Not presented are the results of tests which repeat the conditions of simulated high-sulfur-coal tests previously conducted on the Arapahoe pilot unit and of actual high-sulfur-coal tests conducted at the full-scale Riverside system. In general, these repeat test results show good agreement with the prior SO_2 removal data. Agreement with the full-scale SO_2 removal data in particular provides confidence in the ability to scale up results from this pilot unit.

Figure 3 focuses on SO_2 removal across the fabric filter. In Figure 3, the actual percent SO_2 removal across the fabric filter is presented as a function of the reagent ratio fed to the spray dryer. The percent removal across the fabric filter here is calculated by dividing the decrease in SO_2 content across the fabric filter by the SO_2 content of the flue gas entering the fabric filter. The actual percent SO_2 removal across the fabric filter ranges from about 20 to 30 percent at the lowest spray dryer reagent ratios to about 70 to 80 percent at the highest reagent ratios.

Although the data in Figure 3 indicate that at some conditions the fabric filter can be very effective in removing the SO_2 remaining in the flue gas that leaves the spray dryer, it does not convey how efficiently the fabric filter utilizes the remaining unreacted lime content in the solids collected there. Figure 4 expresses the percent SO_2 removal in the fabric filter, based on the fabric filter inlet SO_2 concentration, as a function of the estimated amount of lime in the solids collected there.

The amount of unreacted lime in those solids as they enter the fabric filter has to be estimated because this value is not directly measured. The amount of lime added at the spray dryer and the amount which has reacted with flue gas SO_2 there are known, because the SO_2 removal in the spray dryer is measured. Also, a few material balance efforts have indicated that about 30 percent of the solids entering this particular spray dryer vessel drop out in the bottom, so on the average only about 70 percent of the unreacted lime is carried over to the fabric filter to contribute to SO_2 removal there. The estimated reagent ratio at the fabric filter in Figure 4 is then 70 percent of: the number of moles of fresh lime added at the spray dryer minus the moles of lime known to have reacted with SO_2 there, divided by the number of moles of SO_2 remaining in the flue gas as it enters the fabric filter. Other factors that may reduce the amount of unreacted lime that is available to react with flue gas SO_2 in the fabric filter are not accounted for in this estimate. Such reductions might occur because solids drop directly into the fabric filter hoppers and are not deposited on the bags, because part of the unreacted lime is converted to less reactive calcium carbonate by reaction with flue gas CO_2, and because some of the lime particles may be completely encapsulated in a shell of calcium sulfite and calcium sulfate.

Given these limitations on the accuracy of the estimated reagent ratio available at the fabric filter, the data plotted in Figure 4 indicate a good correlation between percent SO_2 removal across the fabric filter and the amount of unreacted lime available there. The amounts of unreacted lime estimated to be available range from about 70 percent of that theoretically required to react with all of the remaining SO_2 at the fabric filter inlet (a 0.7 reagent ratio) to nearly three times the theoretically required amount (a 2.8 reagent ratio). The utilization of the unreacted lime estimated to be available for SO_2 removal in the fabric filter varies from about 40 percent at the lower estimated reagent ratios to less than 30 percent at the highest values. In contrast, the data plotted in Figure 2 indicate about 50 to 60 percent utilization of fresh lime in the spray dryer.

In summary, for these high sulfur conditions, the fabric filter has been found to contribute about 10 to nearly 20 percentage points to the overall system SO_2 removal performance. This corresponds from about 20 to over 80 percent removal of the flue gas SO_2 content entering the fabric filter. While the fabric filter has been found at some conditions to be an efficient SO_2 removal device, it is not very efficient at utilizing the unreacted lime collected there. At the highest removal levels of 80 percent of the fabric filter inlet SO_2 content, less than 30 percent of the unreacted lime estimated to be available at the fabric filter is utilized.

PARTICULATE EMISSIONS CONTROL

The flue gas entering the spray dryer contains a fly ash mass loading of about 2 to 4 grains per dry standard cubic foot (gr/dscf). The particulate mass loading at the outlet of the spray dryer (fabric filter inlet) has been measured to vary from 7 to over 15 gr/dscf, depending on spray dryer operating conditions. Thus, because of the lime and recycle solids introduced in the spray dryer, the mass loading of the flue gas entering the fabric filter is much higher than that of a typical flue gas from a pulverized-coal-fired boiler. The pilot fabric filter has been found to be a very efficient particulate removal device, though. Fabric filter outlet mass loadings have been measured to vary from 0.0002 to 0.0013 gr/dscf during normal operation downstream of the spray dryer. These levels correspond to removal efficiencies of 99.98 to over 99.99. Furthermore, these outlet levels are at least an order of magnitude below current New Source Performance Standards requirements of 0.03 $lb/10^6$ Btu (3).

Opacity measurements at the fabric filter outlet indicate values below 2 percent. Actually, the values measured likely represent the accuracy of the zero calibration of the opacity monitor rather than any significant opacity of the fabric filter outlet gas. Cleaning spikes, when a just-cleaned compartment is put back in service, raise the opacity readings above baseline values by 0.5 to 1 percent for several seconds. Also, when fabric filter upsets such as a torn bag or a bag loose from a thimble occur, the fabric filter outlet opacity can increase to 10 percent or greater.

FABRIC FILTER PRESSURE DROP

Before presenting fabric filter pressure drop data, it is worthwhile to discuss its operating history since initial start-up. The pilot fabric filter was put into flue gas service in April, 1987. A pressure-initiated cleaning sequence was employed, with cleaning initiated at a flange-to-flange pressure drop of 4 in. H_2O. Pressure-initiated cleaning was employed at first in order to allow a residual dust cake layer to be established on the bags. It took nearly a day before the pressure drop reached the level which initiated a cleaning. After about two days of operation on normal, 300°F flue gas, slurry feed to the spray dryer upstream of the fabric filter began. During the remainder of April, the fabric filter was operated on a pressure-initiated cleaning frequency, and the spray dryer upstream operated intermittently. After a power plant outage in early May, continuous spray dryer operation began. From mid-May to mid-August, the pilot fabric filter operated about 1700 hours, mostly with the upstream spray dryer in service, and went through approximately 15 start-up and shut down cycles. The fabric filter operated at an air-to-cloth ratio of about 1.2 to 1.4 acfm/ft^2, and was cleaned on a time- or pressure-drop-initiated cycle. The fabric filter was cleaned every three hours, or sooner if the flange-to-flange pressure drop increased to 5 in. H_2O. When cleaning was initiated, all five compartments cleaned in succession. Actual time between cleaning varied from a minimum of about one hour to the full three hours.

In early August, the mode of fabric filter operation was revised to more closely reflect utility fabric filter practice. The number of bags in each compartment was reduced to raise the air-to-cloth ratio to the current value of about 1.6 to 1.7 acfm/ft^2; the cleaning program was revised for time-only initiation (with a 90-minute time between cleaning); and the cleaning program was changed from batch cleaning. That is, the cleaning program was changed so that a different compartment was cleaned every 18 minutes. Accordingly, 90 minutes were required to clean all five compartments.

Figure 5 summarizes the pilot unit flange-to-flange pressure drop performance from the middle of August, 1987 through the middle of February, 1988. These data represent over 3000 hours of fabric filter operation. Because all previous pressure drop data were at other operating conditions, they are not included with the data plotted here. The periods where data are missing from Figure 5 represent times when the spray dryer and/or fabric filter were not in service. Over this period, the fabric filter was started up and shut down approximately two dozen times. All of the data in Figure 5 represent operation at an air-to-cloth ratio of approximately 1.6 to 1.7 acfm/ft^2.

The data in Figure 5 average about 4.5 to 5.0 in. H_2O for a flange-to-flange pressure loss at the beginning of the period, then about 6.0 in. H_2O for the latest month. The first, temporary increase in the average value occurs during the month of September and corresponds to a period where the reverse-gas flow rate was inadvertently reduced. This rather site-specific problem is not discussed in this paper, but it has caused the reverse-gas flow rate on occasion to drop well below the desired 2260 acfm flow rate which produces a reverse-gas air-to-cloth ratio of 1.5 acfm/ft^2. A smaller, permanent increase from below 5 in. H_2O to just above that value occurs at the end of October. At that time orifices were installed at the outlet of each compartment to allow measurement of the flue gas flow rate through individual compartments. These orifices produce a permanent pressure loss of about 0.5 in. H_2O. A rapid increase from an average between 5.0 and 5.5 in. H_2O to over 6.0 in. H_2O in December also appears to be the result of problems in the reverse-gas system that are mentioned above.

In general, there is a slight upward trend in the pressure drop values. However, these pressure drop values are within the range of typical pressure drop performance for conventional utility reverse-gas fabric filters (4). This indicates that there has been no significant adverse effect of the spray dryer on fabric filter pressure drop performance. However, it should be noted that these data represent a relatively short period of fabric filter operation (i.e. just over 3000 hours), and reflect numerous start-ups and shut downs as well as frequent changes in spray dryer conditions while short-term (three-day) SO_2 removal tests are being conducted.

FABRIC FILTER BAG WEIGHTS

Bag weight measurements have recently been made in all five compartments. Six of the 33 bags in service in each compartment were weighed, from the same positions within each compartment. The average bag weight gains (above the weight of a new bag) are summarized below:

Compartment	Average Bag Weight Gain	Equivalent Areal Loading
A	20 lb	0.44 lb/ft^2
B	17 lb	0.37 lb/ft^2
C	20 lb	0.44 lb/ft^2
D	20 lb	0.44 lb/ft^2
E	18 lb	0.39 lb/ft^2

The equivalent areal loadings represent an average based on weight gain per bag and total bag filtering area. These bag weight data suggest that filtering properties are not different from compartment to compartment. The variations in average bag weight gain from compartment to compartment appear to be within the normal variability of such measurements. The average areal loading values are slightly higher than values measured after long-term operation of the fabric filter compartment downstream of the pilot spray dryer at Arapahoe, but only about one-half of steady-state values measured for full-scale reverse-gas fabric filters in high sulfur coal service. Of course, these bag weight data represent the first time that weight gain measurements have been made for this pilot fabric filter. Future measurements will have to be made after continued operation to determine whether bag weights and equivalent areal loadings have stabilized, or are still increasing.

FABRIC FILTER OPERATING EXPERIENCE

In addition to determining quantifiable performance characteristics of a fabric filter downstream of a spray dryer, such as SO_2 removal, particulate removal, and pressure drop, an objective of this pilot study is to determine what other impacts the spray dryer might have on fabric filter operation. The following is a brief discussion of the impacts that have been observed after nearly 10 months of operation of the pilot fabric filter downstream of the spray dryer. A number of operating problems are described, ranging from relatively minor to more serious in nature. Not all appear to be a direct impact of having a spray dryer upstream. Those which appear to be unrelated to spray dryer operation are only briefly discussed.

WALL CORROSION

One of the most significant impacts of the spray dryer on the flue gas entering the fabric filter, other than the obvious increase in mass loading, is that the flue gas is cooled and humidified. For conventional fabric filter applications to utility boilers, the entering flue gas is about 200°F above the flue gas moisture dew point. For a spray dryer application, if the spray dryer is operated at a 20°F approach to adiabatic saturation, the flue gas entering the fabric filter is only about 22°F above the moisture dew point. With any significant cooling, as in areas where the integrity of the insulation is poor or if there are ambient air leaks around a door, fabric filter compartment walls may experience localized condensation of moisture from the flue gas. Once condensation forms, SO_2 (and perhaps SO_3) from the flue gas will absorb into the droplets to equilibrium levels, forming a mixture of sulfurous and sulfuric acid. Similarly, any remaining SO_3 in the flue gas on the downstream side of the filter bags may directly condense as sulfuric acid. Because the compartment walls are on the "clean" side of the bags, there is virtually no lime present to neutralize the acid which forms. This collected acid would be expected to promote corrosion of the mild steel compartment walls.

Significant corrosion was noted in the fabric filter compartment downstream of the pilot spray dryer at Arapahoe (5). There the compartment walls were sandblasted and a number of protective coatings were applied. A relatively inexpensive vinyl ester coating was applied to most of the compartment walls, and more expensive epoxies were applied to portions of the compartment doors. The vinyl ester coating greatly reduced corrosion tendencies, but had lost much of its effectiveness by the time the spray dryer test program was completed approximately two years later. Several of the epoxies applied to the compartment doors were virtually unaffected by the corrosive environment over that two-year period.

Although it was anticipated that corrosion would be experienced in the fabric filter downstream of the spray dryer at the High Sulfur Test Center, it was decided to start out with uncoated walls in order to evaluate the extent of the corrosive tendencies. The fabric filter at Arapahoe was designed to treat normal hot (300°F) flue gas, and so had only 2 inches of insulation installed. The fabric filter at the High Sulfur Test Center is more appropriately insulated to a 6-inch thickness, and hence might have shown reduced tendencies for wall corrosion.

Corrosion has in fact been experienced on the compartment walls of the pilot fabric filter. The corrosion appears as large areas of exfoliating oxidation products, which tend to flake and fall from the walls. In at least one instance, it appears that a large flake of exfoliated material has torn a bag as it fell to the tubesheet. As expected, corrosion is worst near the lower compartment doors, where door seal leaks are obviously occurring. Another area where corrosion is particularly bad is on the north wall of each compartment, near the point where the uninsulated bag shaker mechanisms that were installed for future shake-deflate operation penetrate the compartment walls. The corrosion appears worst in end Compartments A and E.

A small five-compartment fabric filter such as this pilot unit represents a worst case for such corrosion tendencies, in that there is a large amount of wall surface area relative to the volume of flue gas treated. Also, the five-compartment in-line arrangement results in two of the compartments (A and E) having three walls with exterior exposures (i.e., only one wall common to another compartment) and the remaining three have two exterior exposures. The worst corrosion is seen on the exterior walls in this fabric filter, whereas the few interior walls are only lightly corroded. In contrast, a large utility fabric filter with individual casings consisting of two rows of 12 compartments would have only 4 of 24 compartments with two walls having an exterior exposure. The remaining 20 would have only one.

Although pilot installations would be expected to have worse corrosion tendencies than full-scale fabric filters downstream of spray dryers, similar corrosion tendencies have been observed in full-scale spray dryer/fabric filter installations in low-sulfur coal applications (5). One full-scale spray dryer/fabric filter installation currently undergoing start-up has had protective coatings applied to practically all of the fabric filter compartment wall surfaces (6).

To confirm that condensation is indeed the primary cause of the corrosion seen in this pilot-scale installation, each compartment in the pilot fabric filter has had thermocouples installed along the walls to allow monitoring of wall temperatures. After a number of months of monitoring, it is apparent that wall temperatures are on occasion dropping below the flue gas dew point. During one episode where the ambient temperature was 16°F, wall temperatures on the external walls of the compartments were measured at 115°F to 118°F. The flue gas temperature during this period was about 145°F, and the dew point was estimated to be between 120°F and 125°F, so this was clearly a period when condensation was occurring.

Several actions have been taken to slow or eliminate this corrosive action. First, the shaker mechanisms which penetrate the walls of all compartments were removed for this reverse-gas cleaning test period, and the exposed areas where these mechanisms were mounted were insulated. This greatly improved the integrity of the insulation coverage of the compartment walls. Also, improved door seal materials are being tested in an effort to eliminate or at least reduce inleakage of cold air. Finally, protective coatings will likely be effective at minimizing this problem. The entire fabric filter will soon be sandblasted and a number of different protective coatings will applied in order to evaluate their relative effectiveness.

BAG CLAMP FAILURES

In addition to the wall corrosion in the fabric filter, a number of bag clamp failures have been experienced. When the bags were originally installed in the fabric filter in the spring of 1987, the bags were attached to the thimbles on the tubesheet with screw clamps made of a 300 series stainless steel. Within a matter of months, these clamps began to fail, allowing bags to come loose from the tubesheet thimbles. At first it was thought that the clamps were failing due to being overtightened initially. However, after further operation, replacement clamps which were known to have been properly tightened failed as well.

The failures have been attributed to stress corrosion cracking due to the effects of chlorides. Similar failures were observed in pilot testing at the EPRI Arapahoe test facility (7). In this prior instance, stainless steel bag clamps were observed to fail in the fabric filter compartment downstream of a pilot spray dryer during and after tests where calcium chloride was added to the atomizer feed slurry as an SO_2 removal performance additive. A failure analysis of the 300 series stainless steel clamps there indicated that chlorides were responsible for stress corrosion cracking failures. Although the chlorides are theoretically on the clean side of the bags and are present in the solid phase, it appears that enough chloride-containing solids bleed through the bags and that the flue gas humidity is high enough for chloride attack to occur.

While these failures were explained at Arapahoe as having resulted from testing with calcium chloride addition, no such testing has been conducted at the High Sulfur Test Center. However, the coal fired at the Somerset station has a significant chloride content (0.1 wt. percent). The flue gas produced from the Somerset boiler has an HCl content of about 40 to 60 ppmv. This HCl, when absorbed by the spray dryer/fabric filter FGD system, produces a chloride content which typically measures 0.3 to 0.4 weight percent in the solids collected in the fabric filter. This chloride level equals that in the fabric filter solids from the majority of the chloride addition tests conducted at Arapahoe.

It is apparent that screw clamps made of 300 series stainless steel will be unsuitable for use in the fabric filter downstream of a spray dryer which treats flue gas with any appreciable chloride content. A number of unsuccessful efforts have been made to locate suitable screw clamps made of higher alloys or perhaps with a cadmium plating. A more novel solution now undergoing testing is to replace failed stainless steel screw clamps with large Teflon cable ties. Two cable ties linked together will go around the bag on the thimble, and by using a pair of pliers to pull them tight, they sufficiently clamp the bag. It is possible that the bag design used in some fabric filters, where a separate bag clamp is not required to attach the bag to the tubesheet, may be less subject to chloride-related failures. However, the effectiveness of such a design is not to our knowledge demonstrated in long-term operation with significant chloride levels in the solids collected.

Also, it should be noted that since start-up, the residual dust cake weight on the bags has increased. This tends to reduce the upward tension on the bags. Furthermore, where the bags attach to the thimble the cuffs have begun to adhere to the thimbles. This is apparently an effect of a general corrosion of mild steel components within the fabric filter. Because of this adhesion between the cuffs and the thimbles, and the reduced upward tension on the bags, as bag clamps now fail they can go unnoticed because the bags remain on the thimbles in spite of the failed clamps.

SOLIDS HANDLING SYSTEM

The solids collected in the fabric filter are evacuated from the hoppers below each compartment by a vacuum pneumatic conveying system. The compartments are isolated from the vacuum system by air-actuated knife gate valves. These valves frequently require maintenance, because of the buildup of solids on the track and seat for the knife blade. These buildups cause the knife blade to stick, generally in the closed position, or cause the valve to not seal properly. These problems are eliminated by removing the valve and cleaning the valve internals. It is apparently the cohesive nature of the spray-dried solids which make this a particular problem. The solids contain from 0.5 to just less than 1.0 weight percent moisture and almost always contain excess lime. Both of these constituents can make these solids more cohesive than normal fly ash.

Full-scale spray dryer/fabric filter FGD systems, all in low sulfur coal applications, have employed a variety of solids handling system designs. Vacuum pneumatic systems, mechanical systems, and pressurized pneumatic systems (dilute-phase and dense-phase transport) are all used in spray dryer FGD applications. Most of these spray dryer FGD systems report solids handling system problems (8). Based on the experiences at this pilot unit, it appears that solids handling will be troublesome in high-sulfur-coal applications of spray dryer FGD technology as well. EPRI plans a future research effort to develop an optimum solids handling system design that would minimize solids handling problems for these FGD systems.

FUTURE TESTING

Current research on the spray dryer/fabric filter pilot unit is oriented towards short-term SO_2 removal tests. These tests are scheduled to continue through May of 1988. A goal of these tests is to define the capabilities of the process over a wide range of operating conditions and to explore process improvements that may reduce SO_2 control costs as wells as expand the applicability of the process to coals with higher sulfur levels.

Future testing will include long-term (i.e., several month) tests aimed at characterizing and optimizing fabric filter operation downstream of a high sulfur spray dryer. Long-term tests will likely explore the impacts of varied air-to-cloth ratio, filtering dwell, and method of cleaning (reverse-gas, reverse-gas with sonic assistance, or shake-deflate cleaning). Also, bag fabric comparisons will be conducted, with two to three of the five compartments to be bagged with alternate fabrics.

Other particulate control-related testing will include evaluation of a spray dryer FGD system with a pulse-jet fabric filter, and limited testing with an ESP particulate control device.

SUMMARY

After nearly 10 months of operation of the spray dryer/fabric filter FGD pilot unit, the following observations may be made:

- The fabric filter contributes significantly to overall system SO_2 removal performance. The fabric filter contributes 10 to 20 percentage points to system SO_2 removal at overall removal levels of 70 to over 90 percent. With the contribution of the fabric filter to overall removal, it appears that spray dryer/fabric filter FGD systems are capable of meeting the 90 percent SO_2 level required by New Source Performance Standards for systems firing high sulfur coals. These observations are based upon spray dryer inlet SO_2 levels of 2500 ppmv or less and inlet temperatures of 300°F or greater.

  The fabric filter has been observed to remove 20 to over 80 percent of the SO_2 in the flue gas at the fabric filter entrance. However, the fabric filter does not efficiently utilize the unreacted lime in the entering spray-dried solids. It has been estimated that generally less than half of the unreacted lime in the spray-dried solids entering the fabric filter is utilized to remove flue gas SO_2 there.

- The spray dryer upstream has been observed to increase the solids loading of the flue gas entering the fabric filter from 2 to 4 gr/dscf to 7 to over 15 gr/dscf. In spite of this increase in inlet mass loading, fabric filter outlet emissions have remained quite low, well below the NSPS-mandated level of 0.03 $lb/10^6$ Btu. Except during very short-term cleaning spikes and during episodes of torn or loose bags, fabric filter outlet opacity has remained at negligible levels.

- After just over 5000 hours of fabric filter operation, pressure drop values do not appear to be adversely affected by the spray dryer upstream, nor by the numerous (approximately 40) start-ups and shut downs. Flange-to-flange pressure drop values average about 6 in. H_2O. However, one component of this pressure drop value is the site-specific entrance and exit loss. A more appropriate comparison value is the tubesheet pressure drop, which averages about 4.5 in. H_2O. Both the flange-to-flange and tubesheet pressure drop values are within the normal range of utility fabric filter experience for locations without a spray dryer FGD system.

- Also after 5000 hours of operation, bag weight gain measurements indicate values that are quite acceptable.

- There do appear to be some adverse effects of the spray dryer on the operation of the fabric filter. Operation of the fabric filter treating cool, humidified flue gas makes it susceptible to condensation of flue gas moisture at cold spots on the compartment walls. This moisture condensation leads to a very pronounced tendency for general corrosion of the compartment walls and exfoliation of corrosion byproducts. Full-scale installations of spray dryer/fabric filter FGD systems in high sulfur coal applications will likely have to apply protective coatings to compartment walls. Similar experiences have been noted in low sulfur systems that operate at a relatively close approach to adiabatic saturation.

  Further, if the flue gas treated contains a significant chloride content (e.g., 40 to 60 ppmv as HCl), any 300 series stainless steel components in the fabric filter may be subject to stress corrosion cracking failure. Full-scale installations treating flue gas which contains HCl may have to use bag clamp materials that are resistant to chloride-induced stress corrosion cracking, or possibly employ a bag attaching approach that does not require a separate, stressed bag clamp.

Another effect is that the moist (0.5 to 1.0 percent moisture), lime-containing solids collected in the fabric filter appear to be more cohesive than regular fly ash and may be more likely to stick to and foul solids handling equipment. Similar observations have been made at a number full-scale spray dryer/fabric filter FGD systems in low-sulfur-coal-fired applications. It is apparent that the solids handling systems supplied for these FGD systems must be designed with due consideration for these tendencies, regardless of the coal sulfur content. Development of an optimum solids handling system design for spray dryer FGD systems will likely be the subject of a future EPRI research effort.

To summarize, the results are promising for the applicability of spray dryer/fabric filter FGD systems to high-sulfur-coal applications. SO_2 removal levels of greater than 90 percent and excellent particulate removal results have been measured. Fabric filter pressure drop and bag weight values are, so far, at acceptable levels. A few adverse effects of the spray dryer on the operation of the fabric filter have been noted. However, it appears that these effects could be minimized by proper system design.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

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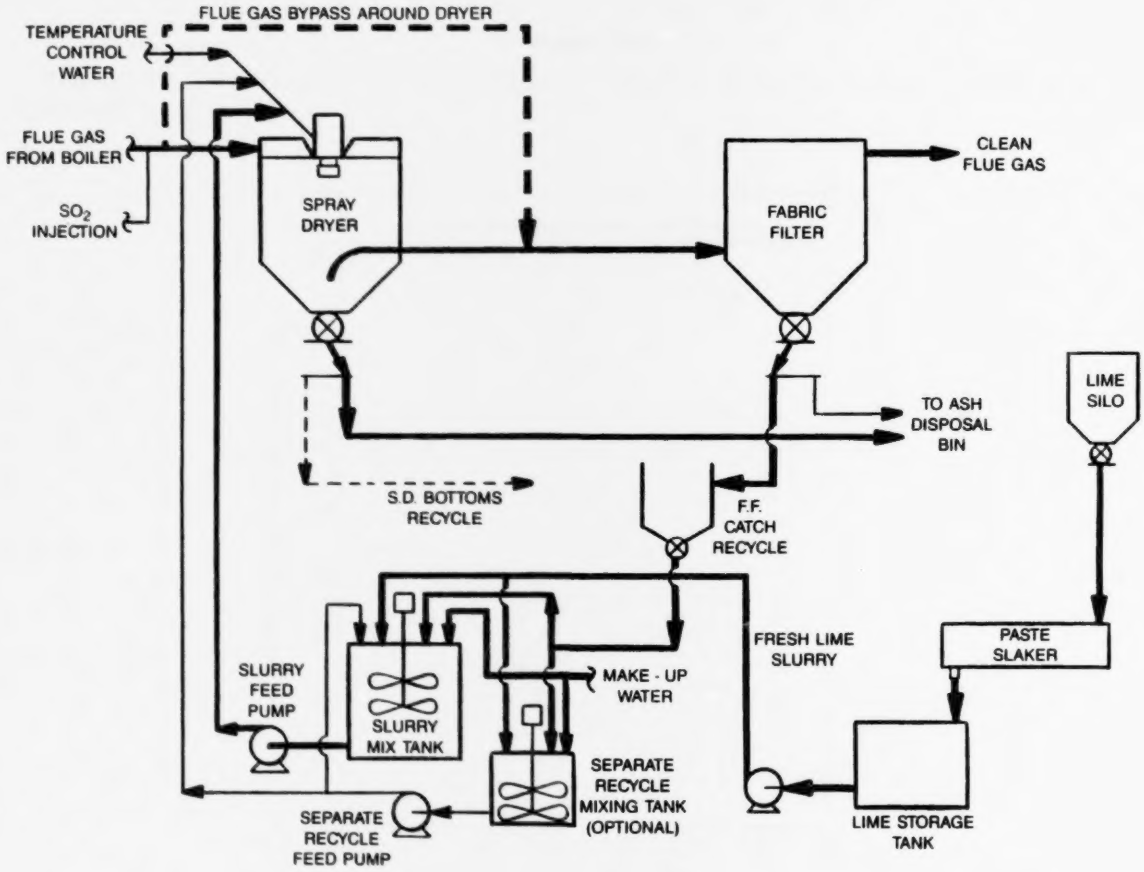

Figure 1. Simplified Process Flow Diagram for the HSTC Spray Dryer/Fabric Filter Pilot Unit

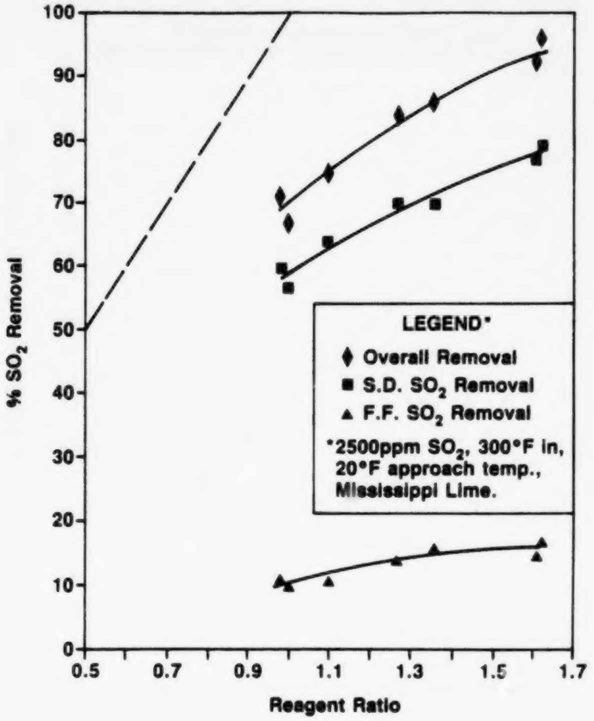

Figure 2. SO$_2$ Removal as a Function of Reagent Ratio with Mississippi Lime Reagent

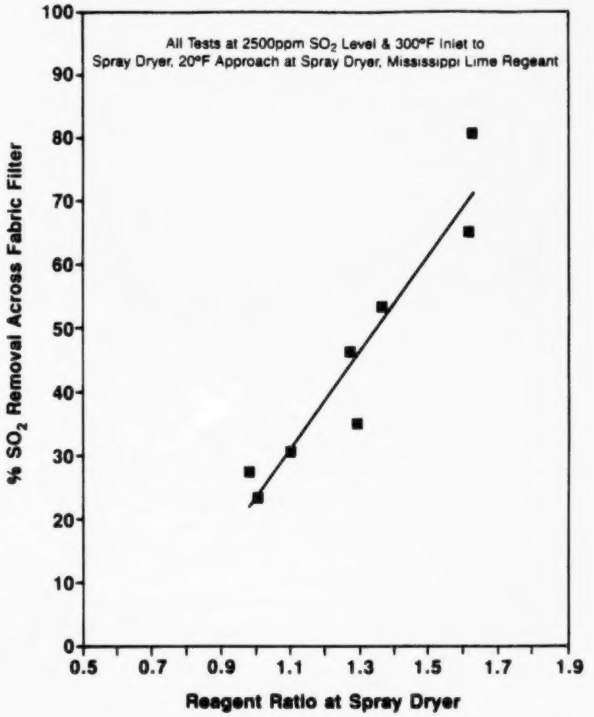

Figure 3. Percent SO$_2$ Removal Across the Fabric Filter as a Function of Reagent Ratio at the Spray Dryer

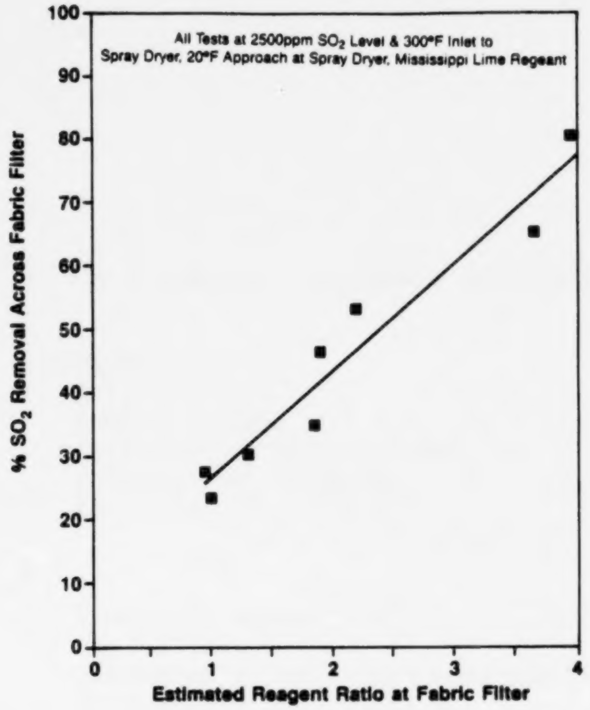

Figure 4. Percent SO_2 Removal Across the Fabric Filter as a Function of the Estimated Reagent Ratio Available There

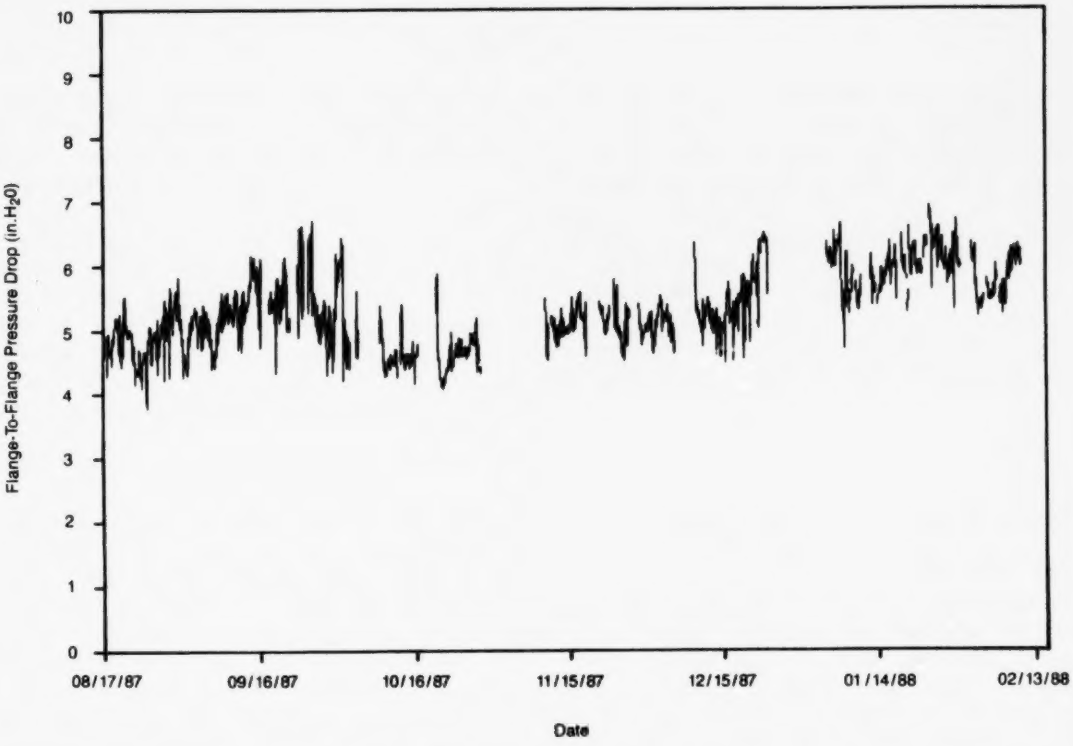

Figure 5. Flange-to-Flange Pressure Drop Data for the HSTC Pilot Fabric Filter, for Operation Downstream of the Pilot Spray Dryer (All at an Air-to-Cloth Ratio of 1.6 to 1.7 acfm/ft^2)

PARTICULATE CONTROL FOR A COMBINED LIME/AMMONIA SPRAY DRYER FOR SO_2 REMOVAL

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ABSTRACT

This paper presents the results of a test program conducted on a 1000 ACFM Spray dryer/baghouse to demonstrate the feasibility of using a conventional $Ca(OH)_2$ - based spray dryer with supplemental injection of gaseous ammonia (NH_3) for SO_2 removal on high sulfur coals. This mode of operation of a spray dryer allows the use of $Ca(OH)_2$ up to a stoichiometric ratio of 0.5 - 0.8, at which point utilization decreases. The NH_3 is injected as a second additive to obtain the overall SO_2 removal efficiency required. This paper will present the results of the initial phase of the study in which a matrix of tests was conducted. Data will be presented on 1) the SO_2 removal efficiency versus stoichiometric ratio 2) the characterization of the reaction products formed, including particle size distributions of the Ca and NH_3 based reacted products, and 3) the effect of the product on the pressure drop of the baghouse.

INTRODUCTION

In the last fifteen years, the spray dryer has been used as a viable control device for SO_2 control on coal-fired boilers with increasing frequency. In general, the spray dryer has used a calcium hydroxide slurry, and has been followed by a fabric filter or electrostatic precipitator for final particulate control. These control systems have generally been limited to medium to low sulfur coals with sulfur contents of less than approximately two percent. For coals with higher sulfur contents, applications of spray dryers have been limited due to the inability to meet stoichiometric requirements as a result of limits on the amount of water which could be atomized and evaporated. In theory, the amount of water available for use in the slurry is limited to the amount required to cool the spray dryer to the water dew point in the flue gas. A further limitation of the application to high sulfur coals is that the efficiency of utilization of the calcium hydroxide has been observed to decrease at the higher slurry concentrations required in the higher SO_2 concentration applications. This has lead to the common use of systems which depend on recycle of the partially-reacted dried solids to provide increased utilization.

The objective of the research program described herein was to evaluate the potential of utilizing a spray dryer/fabric filter to achieve an acceptable level of SO_2 removal in high sulfur coal applications with an overall utilization of near 100 percent. To this end, the research program funded jointly by the Ohio Coal Development Office (Columbus, Ohio), the University of Cincinnati (Cincinnati, Ohio), and the University of Tennessee (Knoxville, Tennessee) was established to investigate and determine the feasibility of utilizing a combined calcium hydroxide/ammonia-based system for SO_2 removal. This research effort began in February, 1987 and is continuing at the time of this writing. This paper presents the preliminary results of the findings to date.

BACKGROUND

$Ca(OH)_2$ STUDIES

Numerous studies have been conducted at both full scale and at the pilot scale demonstrating the relationship between the SO_2 removal efficiency and the operating parameters associated with spray dryer/fabric filter systems(1,2,3). In general, these studies have shown that the efficiency increases with increased stoichiometric ratio and with nearer approaches to saturation at the spray dryer outlet. In addition, the fabric filter serves as a dry reactor due to the increased residence time of the dried solids on the filter bags. For purposes of background development, this paper will use data generated at a 1000 ACFM pilot spray dryer/fabric filter facility located at the University of Tennessee (to be described in a later section).

Typical spray dryer/fabric filter performance data for SO_2 removal efficiency versus stoichiometric ratio (SR) of $Ca(OH)_2$ are shown in Figure 1 for concentrations of 2700-3100 ppm of SO_2. The operating conditions of the spray dryer for the data were as follows: an inlet temperature of 300°F, an approach to saturation of 30°F and no recycle. The removal efficiency of the spray dryer was approximately 27% at a SR of 0.45 resulting in a utilization of only 55%. The spray dryer efficiency increased to 48% at a SR of 1.2, however the utilization had decreased to only 40%. The system, consisting of the spray dryer and fabric filter collector, had an overall utilization of 82% and 48% at SRs of .45 and 1.2, respectively. While utilization improved with the system, it is important to note that without recycle the efficiency of the system is far less than the 90+% that is typically required for a high sulfur coal application. Since the slurry concentration at a SR of 1.2 is approaching 30%, it is evident that the system cannot achieve the required efficiency.

NH_3 STUDIES

In an effort to improve the utilization of the $Ca(OH)_2$-based spray dryer, it was hypothesized that if the system was operated at a SR where the utilization was near 100%, and if a suitable additional reactant could be identified to be used in a supplemental mode, then one might be able to achieve required efficiencies with improved utilization. Based on data obtained previously in all-dry gaseous NH_3 injection studies, it was proposed that NH_3 be used. Research conducted in 1973 by Shale(4,5) and by the authors(6,7) has shown, under proper conditions of flue gas temperature, that gaseous NH_3 will react with gaseous SO_2 and sublime to a fine particulate solid, resulting in the removal of

SO_2. In those studies it was shown that the NH_3 and SO_2 sublimed to a solid at a temperature of 180-190°F. At flue gas temperatures below 180°F, a fabric filter effectively removed the reacted products. At temperatures above the critical value, a visible plume occurred, consisting of sub-micron particles, presumed to be comprised of ammonium sulfite/sulfate and/or ammonium bisulfite/bisulfates(6).

As part of the current study, tests were conducted in the 1000 ACFM pilot plant in which the temperature of the system was controlled by water injection (no calcium hydroxide) to achieve a spray dryer outlet temperature of 155°F (a 30°F approach to saturation). The operating conditions were the same as those discussed in Figure 1. Figure 2 shows a graph of SO_2 removal efficiency versus stoichiometric ratio of NH_3, assuming that two moles of NH_3 were required to react with one mole of SO_2. Based on this assumption, a system efficiency of 40-50% was achieved at a SR of 0.4-0.5 of NH_3, indicating a near perfect utilization. This reconfirmed the data reported previously(5,6,7).

EXPERIMENTAL APPROACH AND PROCEDURES

APPROACH

Based on historical data and the more recent data reconfirming the performance of both $Ca(OH)_2$ and NH_3 based systems, a test program was developed to evaluate the use of a conventional spray dryer with supplemental injection of gaseous ammonia at an SO_2 concentration of 3000 ppm. The objective was to operate the spray dryer/fabric filter in a conventional manner at a $Ca(OH)_2$ stoichiometric ratio of 0.5-0.8. After establishing the baseline operation, supplemental gaseous NH_3 was injected to enhance the removal efficiency, anticipating that it would not result in a deterioration of the already established spray dryer performance.

DESCRIPTION OF THE PILOT PLANT

The pilot plant consisted of a nominal 1000 ACFM system located on a sidestream of the University of Tennessee Riley stoker-fired boilers. The system shown in Figure 3 consisted of a slaking system, a spray dryer (7 feet in diameter) with a variable speed Stork-Bowen AA-6 Atomizer (17,000 rpm, 6 inch centrifugal atomizer), and a 36 bag fabric filter collector operated at an air/cloth ratio of 2.3/1 or 2.3 fpm. The collector contained 13.6 ounce/yd^2 fiberglass filter bags which were cleaned with a low energy compressed air supply (10-13 psi). For the data collected herein, the collector was operated for 4-6 hours between cleaning in a pressure drop range of 3-6 inches of water.

The system provided for continuous monitoring of the static pressure, temperature, and SO_2 concentrations at the inlet and outlet of the spray dryer and baghouse. Continuously extracted samples of flue gas were transported through heat-traced lines to a TECO Model 800/Model 40 SO_2 analyzer. The approach to saturation was measured at all three ports using a hand-held wetted wick thermometer and a thermocouple.

The concentration of SO_2 was maintained by supplementing the boiler's 400-500 ppm by using SO_2 supplied from liquid storage tanks. Ammonia injection was accomplished by monitoring the ammonia injection rates from a liquid ammonia storage tank. In the testing reported in this study, the ammonia was injected upstream of the spray dryer, but downstream of the SO_2 supplemental injection port and sampling port.

TEST PROCEDURES

All data reported represent individual tests of 45-60 minutes in duration. A single data set consisted of monitoring disk speed, slurry injection rate, SO_2 concentrations at three ports, temperature and wet-bulb temperature at three ports, orifice plate pressure drop (flow rate), available CaO, static pressure at three ports, and the ammonia flow rate.

In those tests in which ammonia was injected, the system was operated in an ammonia starved mode at all times that sampling was conducted. This was mandated by the fact that if unreacted ammonia was present, it would pass through the sintered steel filters located at the entrance to the heat trace lines at each sampling port. This would create potential errors in the sampling as reactions would occur in the lines, and would have caused particulate formation in the analyzer. It was possible to achieve an ammonia-starved condition by assuring that two conditions were met: 1) there was always an excess of SO_2 in the exit flue gas, and 2) assuring that there was no visible plume at the exit. In cases where SO_2 was present in the flue gas, the presence of any excess NH_3 resulted in the formation of a visible plume due to the sub-micron particles that were formed at the stack exit as any excess NH_3 reacted with the SO_2. As a secondary check on the absence of NH_3, Sensidine tubes were used to sample for NH_3. While not a very precise technique, the tubes were found to show the presence of NH_3 each time a visible plume was observed.

EXPERIMENTAL RESULTS

STOICHIOMETRIC REQUIREMENTS

A series of three different supplemental NH_3 tests were conducted. The baseline stoichiometric ratio of $Ca(OH)_2$ was 0.5, 0.66, and 0.97 for the three different series. These data are shown in Figures 4 and 5 for the spray dryer and the system (spray dryer + baghouse), respectively. After reaching a stable baseline operation, supplemental NH_3 injection was conducted.

While the baseline data for $Ca(OH)_2$ in Figure 4 and 5 show a clear decrease in utilization as the stoichiometric ratio was increased, the ammonia injection exhibited the same behavior as shown in Figure 2 earlier. The increase in efficiency as a function of SR of NH_3 showed a 100% utilization within the errors of measurement. Only one test resulted in a visible plume. The test conducted at a combined stoichiometric ratio of 1.35 showed the presence of a visible plume. The data in Figure 5 show that in excess of 90% overall efficiency of SO_2 removal can be achieved at a combined $Ca(OH)_2/NH_3$ stoichiometric ratio of 0.95-1.05 thus achieving the required efficiency needed for high sulfur coal at an overall utilization of 85-95%.

PARTICLE SIZE CHARACTERISTICS

One of the primary concerns associated with the supplemental ammonia injection is the formation of sub-micron particles that might eventually lead to increased pressure drop across the baghouse and or deterioration of particulate removal efficiency. During the test program, a series of cascade impactor analyses were conducted using a Pilat Mark 3 cascade impactor. Samples were isokinetically extracted from the spray dryer outlet just upstream of the baghouse for three different conditions: 1) system operating with no injection (only flyash from the boiler was present), 2) system operating with $Ca(OH)_2$ (no NH_3 injection), and 3) system operating with NH3 (no $Ca(OH)_2$ injection).

These data are summarized in the log-probability plots shown in Figure 6. The flyash was found to have an aerodynamic mean diameter (d_{50}) of 1.5-2.0 microns with a geometric standard deviation (σ_g) of greater than 10. The ammonium sulfite/sulfate products had a d_{50} of approximately 0.2 microns and a σ_g of 7.5. The $Ca(OH)_2$ reacted and dried to a $CaSO_3$ product with a d_{50} of 22 microns and a σ_g of 4.6. In the case of these later two analyses, the values were influenced to some extent by the of the flyash which was always present in the boiler flue gas.

PRESSURE DROP

A preliminary analysis of the baghouse pressure drop during the test program showed that the residual pressure drop after cleaning the filter bags was in the range of 3-4 inches of water, depending on the conditions tested. Generally, the baghouse was operated for a 4-6 hour test period and then cleaned. The tests conducted were of such short duration that it is not possible to draw any conclusions regarding the long term effect of the NH_3 particle size on the baghouse pressure drop. Figure 7 does illustrate typical short term pressure drop rise rates for the ammonia-only system, the $Ca(OH)_2$ only system and the combined $Ca(OH)_2/NH_3$ system. Although the NH_3-only system had a slightly higher rise rate (indicative of the smaller particles), it is interesting to note that the combined $Ca(OH)_2/NH_3$ system appears to be similar to the $Ca(OH)_2$ system indicating that the presence of the larger $Ca(OH)_2$ particles provides a buffer, protecting the bags from the sub-micron particles. An evaluation is in progress to develop normalized data to account for differences in the loading to the filter associated with the NH_3-only and $Ca(OH)_2/NH_3$ conditions. Additional testing is needed to determine the long term effect of the combined system.

PRELIMINARY CONCLUSIONS

The above results have shown that it is possible to operate a combined lime/ammonia spray dryer/fabric filter system which can remove SO_2 from high sulfur coals at the pilot plant level. The system was capable of SO_2 removal efficiencies in excess of 90% at stoichiometric ratios of 0.85-1.05. These results are preliminary in that the long term effects of the combined system on the baghouse pressure drop have not been evaluated.

The system was operated at the above conditions without a visible plume. Additional tests are needed to quantify the limits of operation without a visible plume.

In addition to the above findings, there are unanswered questions regarding the nature of the waste products which consist of a combination of calcium and ammonium sulfites/sulfates. Research is currently being conducted to evaluate the feasibility of regenerating the ammonia by slurrying the spray-dried and sublimed waste products, followed by calcium addition. If this research is successful, it may be possible to minimize the use of ammonia and decrease the potential leachability of the waste products.

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The work described in this paper was not funded by the U.S. Environmental Protection Agency. The contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

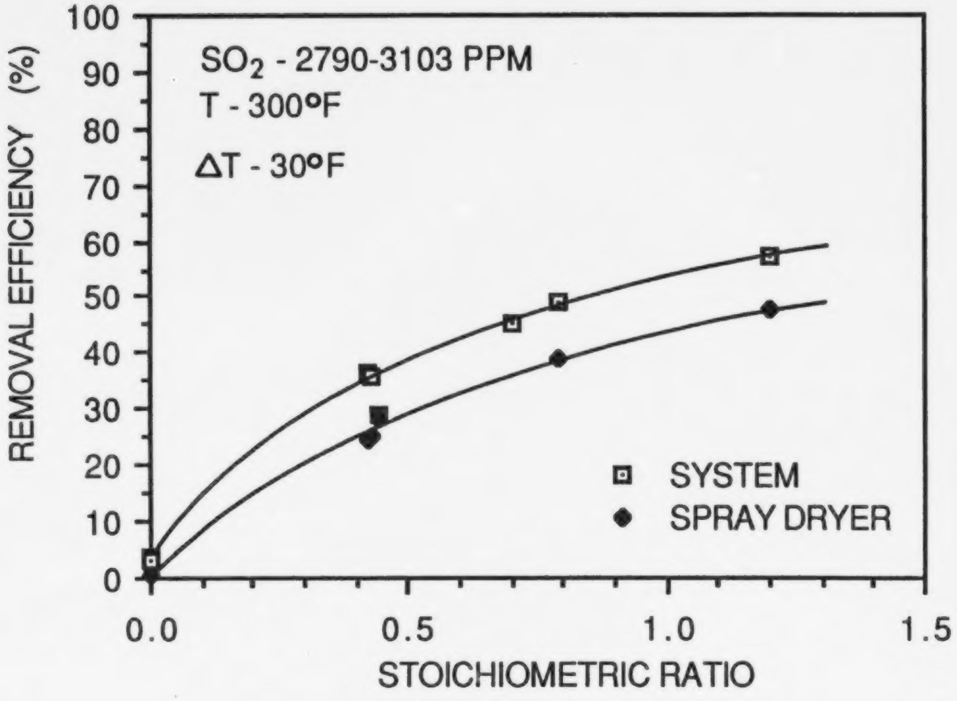

Figure 1. Performance of a $Ca(OH)_2$ Based Spray Dryer/Fabric Filter

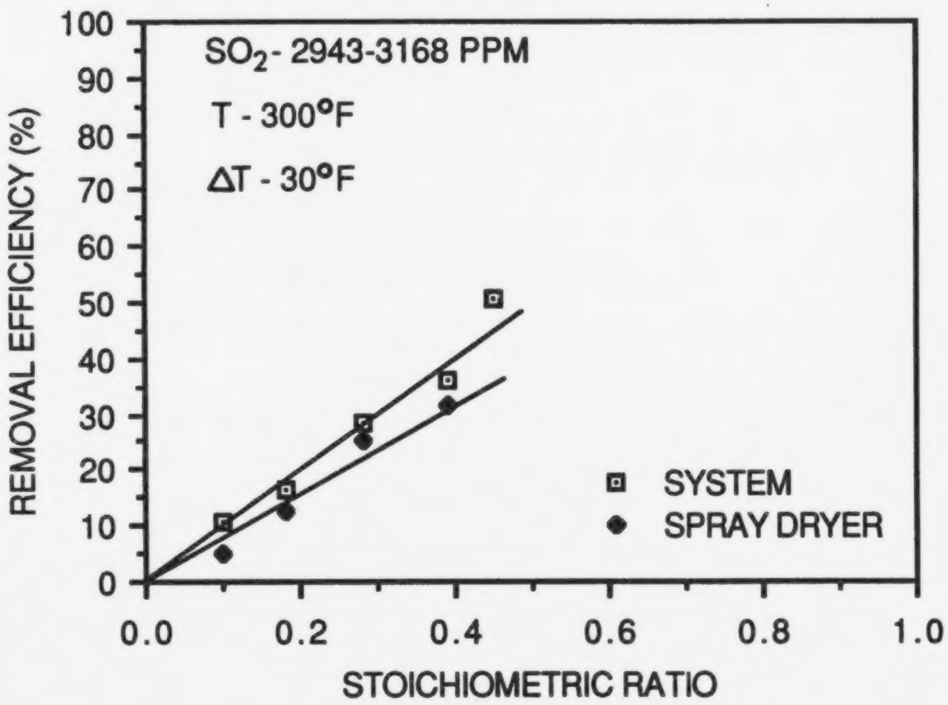

FIGURE 2: Performance of a Gaseous NH_3 Injection System

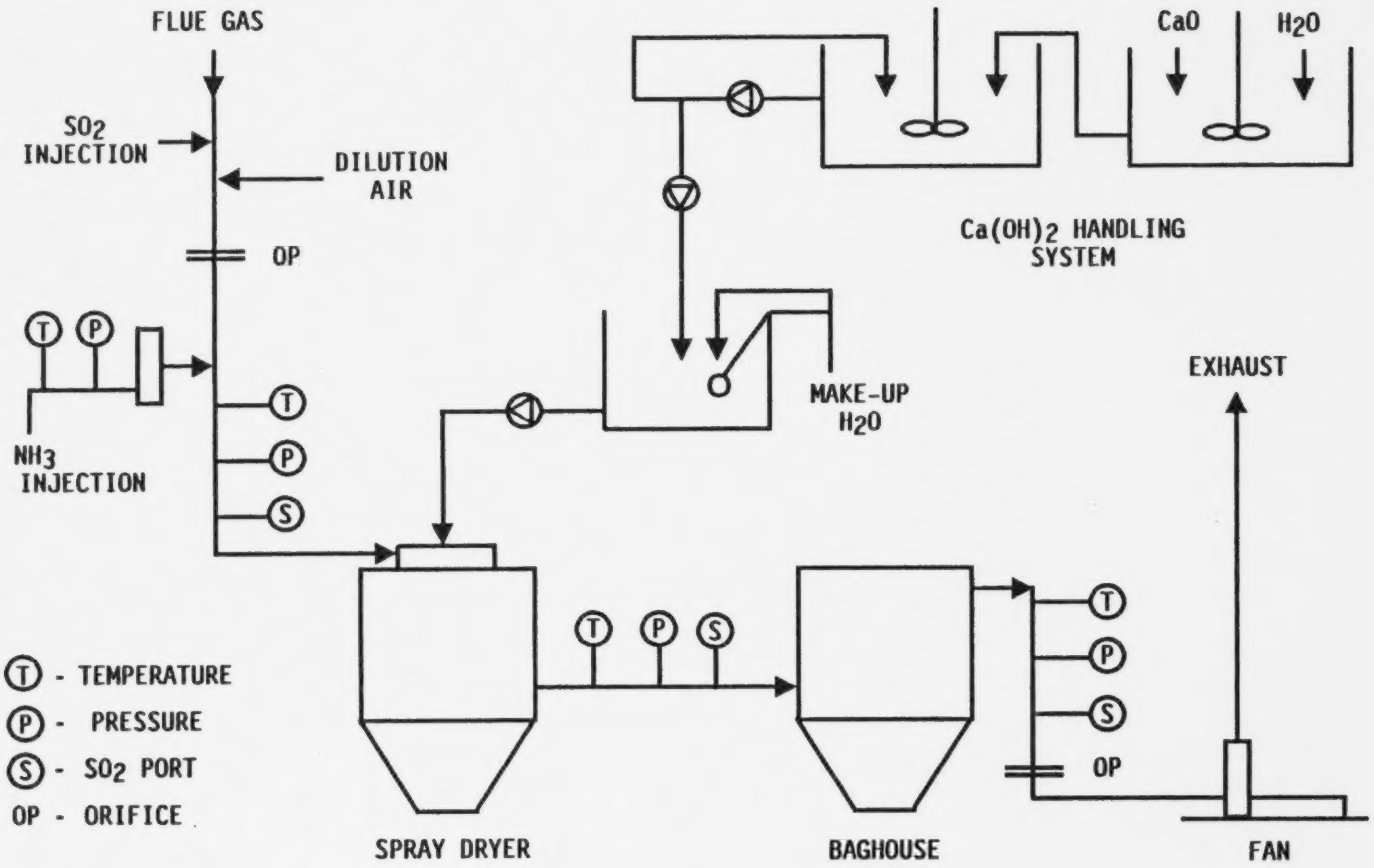

FIGURE 3: Schematic Diagram of the Pilot Spray Dryer/Baghouse Facility

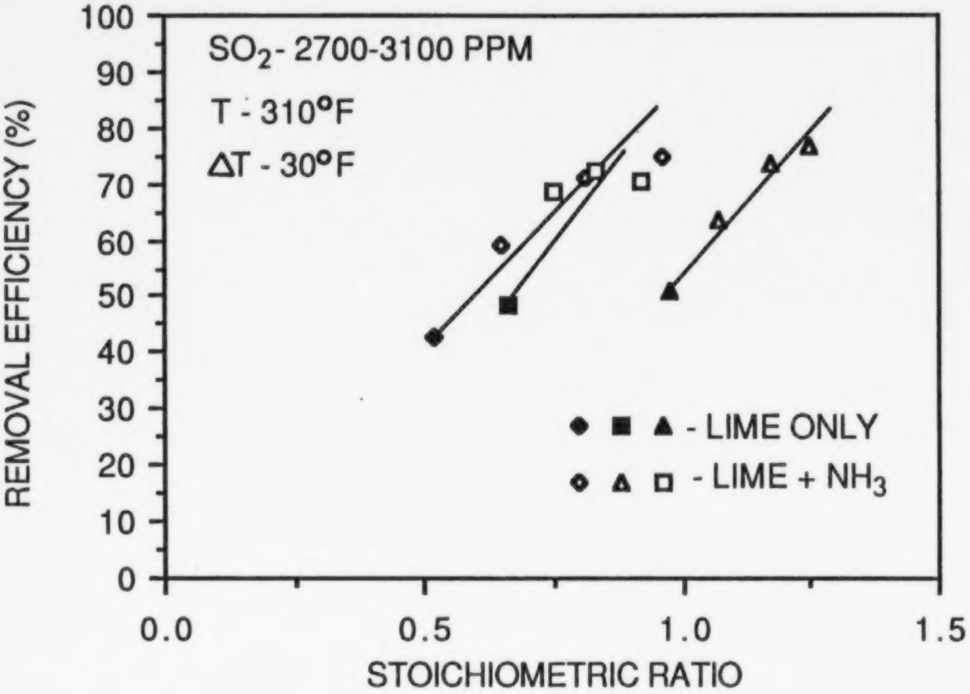

FIGURE 4: Performance of Spray Dryer

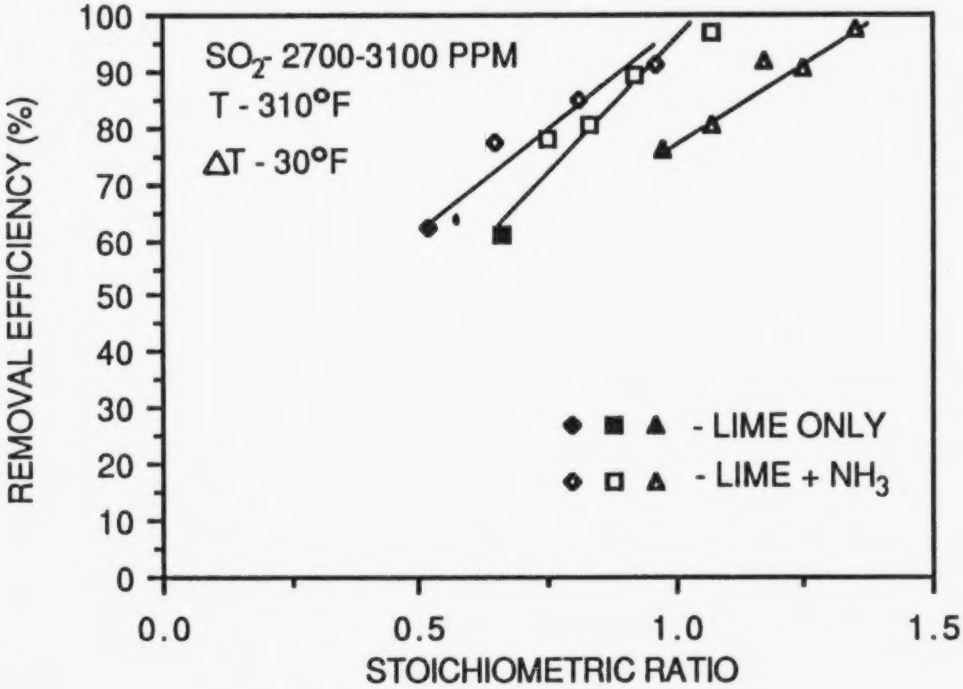

FIGURE 5: Overall Performance of the Spray Dryer/Fabric Filter

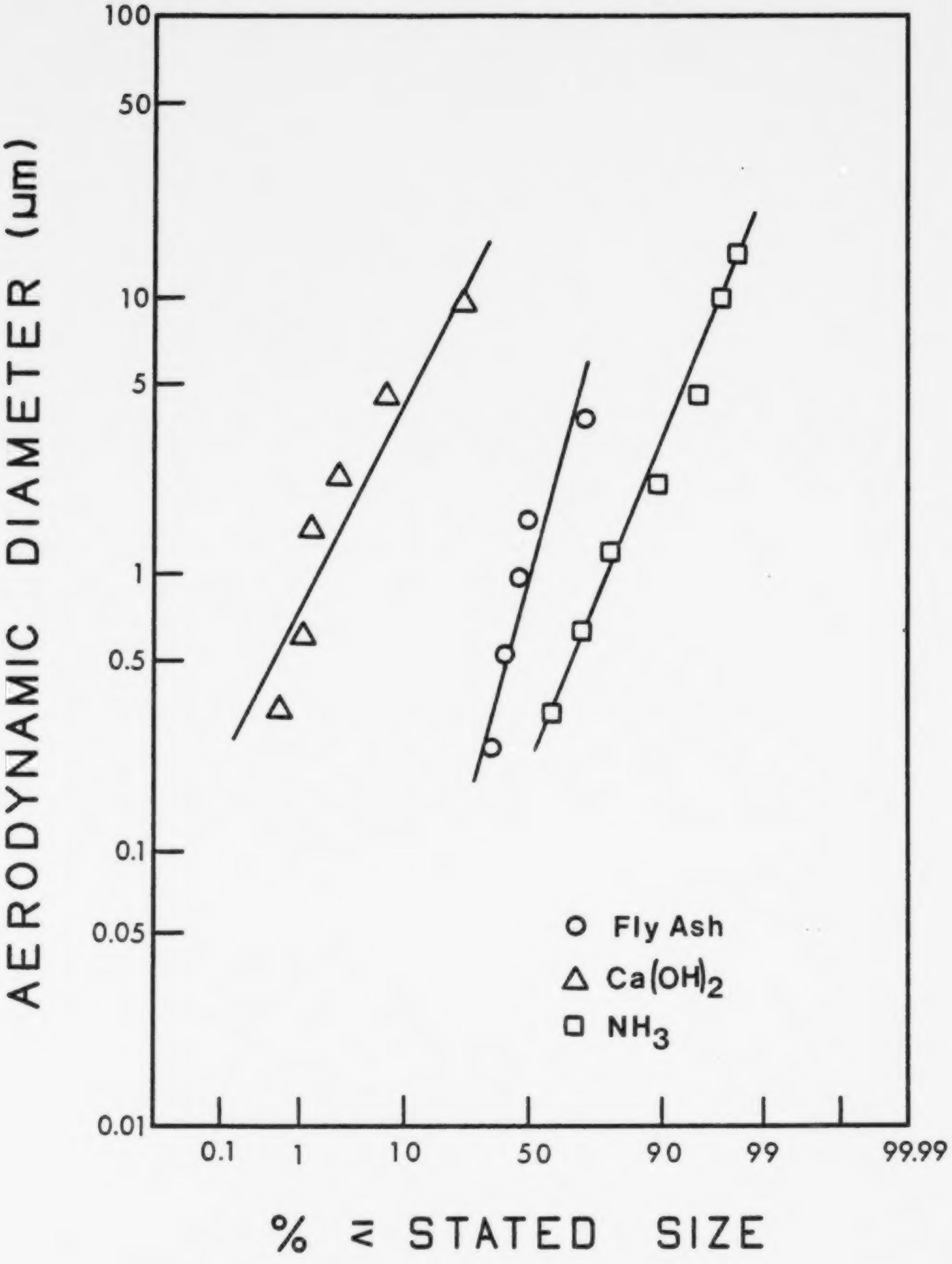

FIGURE 6: Particle Size Distribution for Reacted Products

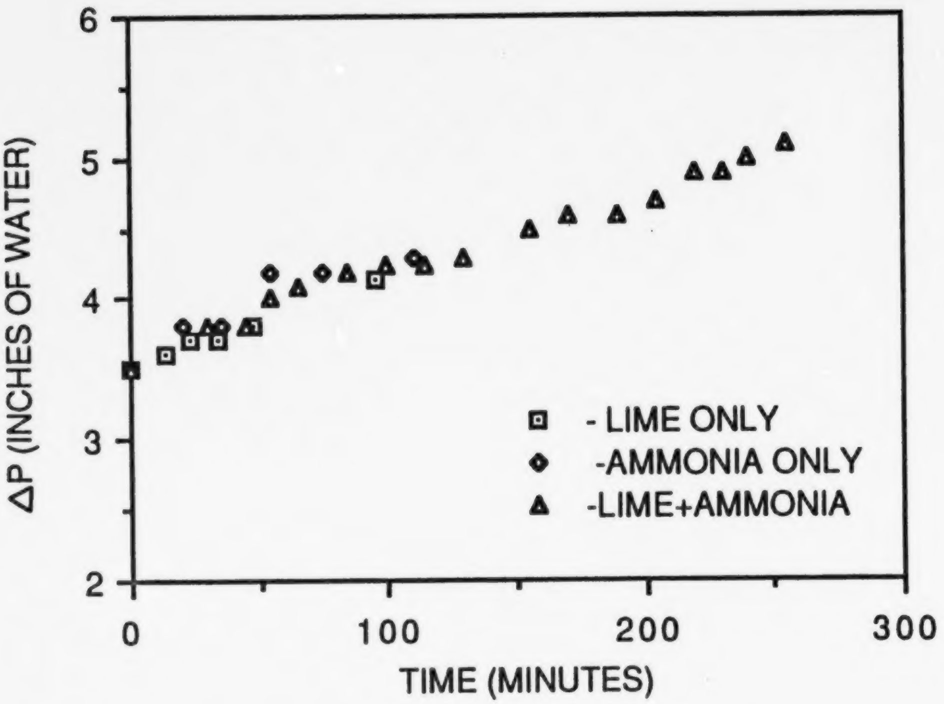

FIGURE 7: Pressure Drop Versus Time for the Baghouse

OPERATING RESULTS OF SEVERAL BAGHOUSES IN DRY FGD SERVICE

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ABSTRACT

Long term operating results on several baghouses operating with dry FGD systems are reviewed. The baghouses are on both industrial and utility- sized projects burning a variety of fuels, with and without product recycle, and are both reverse air and pulse cleaned. Parameters studied are corrosion, bag life, pressure drop, cleaning cycle, maintenance requirements, and emission rates.

I. INTRODUCTION

This paper presents operating experience from baghouses operating as part of dry flue gas desulfurization (F.G.D.) or acid gas control systems in the U.S., Canada and Sweden. The plants chosen for this survey represent a range of dry and wet/dry applications, including utility and industrial sized units utilizing a variety of fuels. Data for both reverse air and pulse cleaned baghouse equipment is included in this paper. A summary of the plants, equipment and applications is given in Table 1.

Data was gathered from each plant location and a review was made of current conditions, operating experience and problems. A description of each system is given in the next section. Baghouse design data is given in Table 2. In evaluating the current condition of each system, typical operating data was gathered, and operational problems discussed. Bag samples were gathered where possible to allow a long term fabric evaluation.

II. SYSTEM DESCRIPTIONS

1. UNIVERSITY OF MINNESOTA

The University of Minnesota Southeast Steam Plant, located in downtown Minnespolis, operates two (2) coal fired boilers that produce steam for campus heating and are capable of driving a 12 MW turbine. Boiler No. 3 is a PC fired Combustion Engineering unit. Boiler No. 4 is a stoker fired Babcock and Wilcox unit. Flue gas from each boiler flows to separate FGD reactors and each reactor is equipped with a single 150 HP rotary atomizer. The gas from each reactor is handled by an eight (8) compartment low ratio, reverse air cleaned baghouse, an ID fan and a separate stack. A vacuum, dilute phase ash handling system is used to evacuate ash collected in the hoppers. Slaked lime is used as the FGD reagent. The collected fly ash and lime is not recycled in this system. Coal sulfur content averages 0.5% and the system operates with 80-95% SO2 removal.

The retrofit FGD system and baghouses were installed in 1982. First firing of Boiler No. 4 occurred in September 1983 and Boiler No. 3 followed in July of 1984. Both baghouses have been in nearly continuous service since that time. The baghouses handle flue gas treated in the FGD reactors over 90 percent of the time and untreated flue gas for the remaining time. Each baghouse is divided into eight (8) compartments, with each compartment containing 133 bags.

The woven fiberglass bags are finished with Teflon B and are eight (8) inches in diameter. The baghouse was designed for a nominal inlet gas flow of 97,000 ACFM at 160 degrees F. The design produces a net filter speed of approximately 2.5 ft/min.

2. ROCKWELL INTERNATIONAL

The Rockwell International, North American Aircraft Operations Plant in Columbus, Ohio operates up to four (4) Riley Stoker boilers to provide plant heating and process steam. The flue gas from the boilers flows into a common duct and is treated in one FGD reactor, equipped with a single 300 HP rotary atomizer. The treated flue gas then flows to an eight (8) compartment, low ratio, reverse air cleaned baghouse, dual ID fans and a single stack. Slaked lime is used as the FGD reagent, without recycle. Coal sulfur range is 0.8 - 3.0%. Due to the fact that this unit operates with only an emissions limit (percent removal is not specified), the system operates with 50-85% SO2 removal.

The retrofit FGD system and baghouse was installed in 1984-1985. First firing of the boilers occurred in 1985 with the scrubber start-up in December, 1985. The baghouse has been in nearly continuous service since that time. Each baghouse compartment contains 149 bags, eight (8) inches in diameter. The bags are woven fiberglass with Teflon B finish. The baghouse was designed for a nominal inlet gas flow of 105,000 ACFM at 170 degrees F. The design produces a net filter speed of approximately 2.1 ft/min.

3. NORTH VALMY

The North Valmy Station, North Valmy, Nevada is jointly owned by Sierra Pacific Power Company and Idaho Power Company. The site has two (2) generating units, each rated at a nominal 250 MW. Unit No. 1 is equipped with a ten (10) compartment low ratio baghouse only (no scrubber). Flue gas from Unit No. 2 is treated in three (3) parallel FGD reactors, each equipped with three (3) 300 HP rotary atomizers. Treated flue gas flows through a common duct to a ten (10) compartment baghouse. Slaked lime is used as the FGD reagent, with recycle of ash and reactor product. The coal sulfur content for both units ranges from 0.4 to 0.7%. The Unit #2 scrubber normally removes 70% of the SO2.

The Unit No. 1 baghouse was installed in 1978-1980 and placed into service in 1980. The Unit No. 2 scrubber and baghouse were constructed in 1982-1984, with start-up beginning in April, 1985. Both units have been in nearly continuous service since their respective start-ups.

Each structural baghouse is divided into ten (10) compartments. Each compartment of Unit No. 2 contains 612 bags, 12 inches in diameter. The Unit No. 1 baghouse contains 648 bags per compartment to accommodate the gas volume required for filtering untreated flue gas. The bags are woven fiberglass with Teflon B finish. Both units were designed for a nominal 1,200,000+ ACFM direct from the air pre-heaters at 260-300 degrees F. The design produces a net filter speed of 2.3 ft/min. Unit No. 2 baghouse was additionally designed for normal operation after the scrubber with 970,000 ACFM at 160 degrees F, producing a net filter speed of 2.0 ft/min.

4. OGDEN/BRISTOL

Ogden Martin Systems of Bristol, Inc. is the owner and operator of a resource recovery facility burning municipal waste from the Bristol, Connecticut area. Two (2) Martin inclined, reverse reciprocating stokers with integrated boilers, supply flue gas to two (2) separate gas cleaning trains. The facility burns 650 tons per day of waste and generates 13 MW of power. Each gas train consists of an acid gas reactor equipped with three (3) dual fluid spray nozzles followed by a four (4) compartment, pulse cleaned baghouse, utilizing off-line cleaning. The term "acid gas control system" is more appropriate for incinerators than "FGD system" since not only SO2 but also HF and HCl are reacted and removed in these systems. Slaked lime, without recycle, is used as the reagent. Acid gas removal exceeds 90% for HCl and 80% for SO2.

The acid gas control system was installed in 1986-1987. Start- up occurred in October, 1987. The baghouses were subject to daily on-off cycling service for a short time. Since mid-December, 1987, both trains have been in nearly continuous service. Each baghouse consists of four (4) modules, with each module containing 266 bags. The bags are woven fiberglass, with Teflon B finish, six (6) inches in diameter. Design of the unit was based on a net filter speed of 3.7 ft/min.

5. QUEBEC CITY

The Quebec City Plant is a municipal solid waste incinerator. A portion of the flue gas from this incinerator was run through a pilot plant equipped with a wet/dry acid gas reactor with one dual fluid atomizer nozzle, a dry injection reactor, and a high ratio, pulse-cleaned baghouse. The wet/dry reactor was used for both direct spray drying of reagents and for humidification/temperature control when the dry reactor was in operation. The pilot plant was used to develop acid gas control performance data and fabric filter emissions control data. Various reagents, including commerical hydrated lime and slaked lime were used. (1)

The pilot plant was installed and operated in 1986. The baghouse was a single module, high ratio type, with six (6) rows of 12 bags/each. The bags were five (5) inches in diameter. Initial bag fabric was Teflon felt with Gortex scrim. Woven glass, double beam construction, with 10 percent Teflon B finish was also tested. Design of the pilot plant was based on a nominal 4,500 ACFM, giving a design filter speed of 4.9 ft/min. The fabric filter handled a range of operating conditions, with both on-line and off-line cleaning used.

6. CITY OF SODERTALJE

The City of Sodertalje, Sweden (near Stockholm) operates three (3) coal fired boilers for district heating. Each unit is rated at 120 MW thermal output. The boilers fire coal from a variety of sources, with a sulfur content of 0.7 to 1.5 %. Each of three (3) independent gas trains consists of an electrostatic precipitator (used as a pre-collector, allowing the sale of flyash), followed by an FGD reactor and baghouse. The FGD reactor uses three (3) dual fluid nozzle atomizers. The gas from each reactor flows to a four (4) chamber, pulse-cleaned baghouse. Lime is used as the FGD reagent, with recycle of ash and reactor product.

The FGD system was installed in 1981-1982 and began operation in October, 1982. The system typically has been operating 5000 hours per year since that time. Each chamber of the baghouse has 540 bags of acid resistant polyamide (Nomex). Each bag is 5 inches in diameter. Design of each unit is based on a nominal gas flow of 129,300 ACFM at a baghouse inlet temperature of 165 degrees F. With on-line cleaning, the design results in a net filter speed of 4.6 ft/min.

7. SYSAV

The SYSAV Plant, located in Malmo, Sweden, is an incinerator burning a combination of municipal and industrial waste. The two (2) moving grate boilers supply 50 MW (thermal) of hot water to the City of Malmo central heating system. Each boiler discharges to a mechanical pre-collector (cyclones), a waste heat boiler, and then to individual dry lime reactors. In the reactors, hydrated lime (calcium hydroxide) is added to the flue gas. The gas from the reactors then passes through a de-energized electrostatic precipitator which acts as a settling chamber and then to a four (4) compartment high ratio baghouse. Removal of HCl exceeds 90%.

The retrofit acid gas control system (dry reactors and baghouses) was installed in 1981. Both boilers are base loaded units that operate more than 8,200 hours per year. Only a one week outage is scheduled each year. Each baghouse compartment contains 308 bags. Each bag is five (5) inches in diameter. Each unit was designed for a nominal inlet flow of 127,000 ACFM at 290-320 degrees F. The design produces a filter speed, with on-line cleaning, of 4.8 ft/min. (2)

A variety of felted bag materials, discussed in Section IV, have been used at this plant.

III. BAGHOUSE OPERATIONAL CHARACTERISTICS

Operational information was collected from all seven (7) plants. All plants except the Quebec City pilot unit are currently in operation. Current operational characteristics and information is summarized in Table 3. The general condition of each unit was evaluated. Plant administrators, operators and maintenance personnel were interviewed to gather operational data and to discuss any problems existing with the units. These observations are presented below.

1. UNIVERSITY OF MINNESOTA

Both baghouses operate extremely well, with an interval of several months between even minor problems. Both units were inspected in June 1987 and were not opened until Unit No. 4 was inspected in January 1988. All of the originally installed bags are intact in each unit. The only problems of a maintenance nature are dust quantity related. The control system is preset to clean whenever the pressure drop across the baghouse reaches 6" W.G. The ash cake on the filter bags is extremely permeable and produces very little flow restriction. In normal operation, it can take 24 to 48 hours to reach the 6" setpoint. Cleaning after such a long filter run can overwhelm the vacuum ash system with ash and plug the hopper outlets. Manually initiating a cleaning cycle one to three (3) times per day avoids this problem. The six (6) inch pressure drop setpoint for cleaning is seldom reached.

Several minor design problems have been addressed. They are all related to cold climate operation (the baghouse damper operators are not enclosed). Boxes with heat lamps have been built around all poppet damper actuators. The seals leak somewhat on the reverse air damper shafts. Tighter seals have been tried, but create new problems by stopping the actuators. Partial enclosures have been added to each end of the baghouse for wind protection. This has helped to reduce the wind-chill effect on both the equipment and operators. A clean gas recirculation bypass has been added to the reverse-air duct, to help keep the reverse air duct warm. This recirculation prevents "dead" gas from cooling and condensing moisture in the reverse air duct when the dampers are closed.

Compartment No. 8 on the baghouse for Boiler No. 4 (Stoker unit) was recently inspected. Some door corrosion from in-leakage and cold exists on this unit. A rust preventative paint that the University has tried did not appear to help. However, the main surfaces of the doors and frame are still in very good condition. The inside of the compartment was only streaked from moisture in a few places, most noteably below the door frames. The stains had not caused any corrosion or metal loss. All outside walls are in excellent condition. Essentially no other corrosion or moisture problems could be found. The compartment was very clean.

Two (2) bags were removed for testing from Compartment No. 8. They had a very light coat of dust. Adjacent bags were shaken by hand and the fallout observed through the open tubes. At most, only a few hands- full of dust could be removed from each bag. All bags appeared to be in as-new condition (these bags were installed in 1982). Tension was checked on several and found to be at the original value.

The only visible signs of problems were on the poppet actuators. Corrosion, steam, moisture, and ice could be found on all reverse air poppet shafts. All other exterior areas of the baghouses looked very good. The hoppers were scraped out last summer, and no hopper pluggages have occurred since that time.

The baghouse for Boiler No. 3 was not inspected inside. The bags in this unit were reported to maintain a heaver dust-cake than bags in Unit No. 4. Considerable dust-cake was found in all bags in this unit when last inspected. Some can be removed with manual shaking but most cannot. The University is sure this baghouse has operated many times at high moisture and low temperature conditions, due to boiler tube leaks. Considerable bag blinding is present in this unit. However, due to the low gas flow that it handles (boiler is de-rated to a single pulverizer) no operational problems have resulted. Essentially, what is now an over-sized baghouse is still quite satisfactory in terms of operating cost with partially blinded bags. Nearly continuous cleaning is required for emergency boiler conditions in which two (2) pulverizers are used. It was reported 110,000 lbs/hr of steam is possible for short time periods, but normal maximum rating is 52,000 lbs/hr. Baghouse door corrosion was reported to be somewhat more extensive in Unit No. 3, but had not progressed or caused any problems.

2. ROCKWELL INTERNATIONAL

The baghouse handles a highly variable gas flow rate due to the cycling on and off of multiple boilers and variable steam requirements. From a process and operational point of view, the baghouse works very well. Pressure drop is usually low due to the low gas flow handled and excellent dust filtration properties. Dust handling from the baghouse hoppers with a vacuum pneumatic conveying system normally works well, inspite of long filter run times and large surges of ash. Exterior examination of the baghouse reveals that the poppet damper actuators and shafts are in good condition. Very little seal leakage occurs on this unit.

The main area of concern with this baghouse is cold-spot areas with moisture condensation and corrosion. This problem was addressed in 1987 with the application of an interior protective coating. All bags were removed and the baghouse compartment walls, doors and doorframes were blast cleaned and coated. New bags were installed and the unit put back in operation in July, 1987.

Inspection was made and a bag removed from Compartment H in January 1988. The bags appeared to be in excellent condition although many were in need of retensioning. The bag removed had only a light coating of ash. Adjacent bags were shaken and only a small amount of dust could be removed, indicating that bag cleaning is adequate. The area of hopper interior that could be observed was clean.

The protective coating on the door and compartment interior walls was in excellent condition. No peeling or blistering could be found. The exterior corners at the tubesheet were dry and had no corrosion. However, the compartment floor below the door and to both sides of the door showed obvious signs of water. Also, a large number of dust-caked water streaks extended down the wall from the compartment vent duct opening to the floor. The new coating did not seem to be affected by these streaks. The door frame and door interior were also streaked but protected from corrosion by the coating.

The baghouse at Rockwell International currently operates with a 15 degree F temperature loss across the unit. The process conditions and unit size parameters are nearly identical to those at the University of Minnesota. A 5 degree F temperature drop is typical for that unit. It is believed that air inleakage at the compartment doors is the major source of condensation and temperature drop in the Rockwell unit. The door leakage problem is tolerable on the upper level since that area is hot. Due to the relatively low gas flow in this unit, cold air sucked into the lower compartment doors does not mix well with the hot gas passing upward through the bags. Thermocouples put into one compartment and measured 180 degrees F at the top and 105 degrees F at the bottom of that compartment. Wet dust or hopper pluggage due to this cold, moist gas is not a problem. The hot inlet gas and hopper heaters keep the dust flowing. The only the cold zone that exists is at the tubesheet level. The reverse air fan casing and lagging also shows signs of corrosion. Both of these problems can be corrected by proper maintenance, as has been accomplished at other installations.

3. NORTH VALMY

The Unit No. 1 baghouse has been included in this paper due to the fact that it provides a nearly identical non-scrubber application to compare with the experience of the Unit No. 2 baghouse. The operation and maintenance characteristics of both units have many similarities. Both baghouses require very little maintenance attention. The exteriors of both units are clean. The poppet damper shafts and actuators on all but the reverse air dampers are in excellent condition. The reverse air damper shaft seals are worn and the shafts are pitted and rusty, but function without problems. The out-leakage of clean gas around these shafts is minor. Door leakage is also minor. The reverse air fans on both units run at very low vibration levels. Reportedly, the majority of the bag tension springs in both units are totally compressed. Design tension is maintained on the bags, and no noticeable problems have resulted.

The major differences between the two (2) units are the result of fabric condition, dust quantity, characteristics and temperature. The eight (8) year-old bags in the Unit No. 1 baghouse are near the end of their useful life. Some bleed-thru of dust occurs on Unit No. 1, raising the opacity to the 5-6 percent level. The Unit No. 1 bags also suffer from considerable fabric blinding. This increases the pressure drop to the point that continuous cleaning is required at full load. Unit No. 1 maintains about 6.5" W.G. pressure drop (one compartment cleaning). Only a few bags have been replaced to date. It is anticipated that a complete change-out will occur in the near future.

The bags in the Unit No. 2 baghouse, while somewhat newer, seem to be benefitting from the protection given by the scrubber. Much less fabric blinding was evident. The opacity on Unit No. 2 is an order of magnitude better than Unit No. 1, with a repeatable average below 0.5 percent, although readings this low are certainly subject to instrument calibration. The Unit No. 2 baghouse has a pressure drop approximately one (1) inch lower than Unit No. 1, averaging about 5.5 inches with one compartment cleaning. Unit No. 2 handles approximately three (3) times more dust, somewhat more gas on a mass basis and 7.5 percent less volumetric flow due to the lower temperature existing after the scrubber.

Ash handling on the Unit No. 2 baghouse was reported to be a constant problem, especially in cold weather. The dust removal system is a positive pressure design, with pressurized feeders that must vent internal pressure as they fill with ash. The main problem is the closing-off of the feeder vent lines by internal plating of moist ash. If the ash stays in the hopper for any length of time, it will not flow out. Continuous ash evacuation helps to even-out the ash flow and reduce pluggage problems.

Door area corrosion exists on Unit No. 2. In an attempt to correct this problem, blast cleaning to white metal followed by one coat of epoxy was undertaken last year. Poor results were found at the next outage. The epoxy had blistered, cracked and generally failed to protect the base metal. Door sealing is not thought to be the problem. The doors are simply too cold to prevent corrosion and the epoxy did not form a suitable barrier. This type of corrosion does not exist to any noticeable extent on Unit No. 1. It must be remembered that Unit No. 2 operates approximately 30 degrees F above the water dew point, while Unit No. 1 is over 100 degrees F warmer.

4. OGDEN/BRISTOL

The short period of incinerator operation at the Ogden/Bristol plant limits the operating experience at this installation. However, indications are that both units are maintaining very low acid gas and particulate emissions levels. Data from the State of Connecticut Compliance Testing has not yet been released.

5. QUEBEC CITY

This test program was completed in 1986. Despite the normal problems associated with a pilot plant installation, the high ratio fabric filter used during this testing operated with minimal difficulties. Pressure drop through the baghouse was maintained at 5.6 to 6.2 inches during both the wet/dry and dry testing by adjusting the cleaning pulse rate.

6. CITY OF SODERTALJE

The three (3) baghouses at Sodertalje continue to operate very well. The baghouse maintains a constant five (5) inch W.G. pressure loss in spite of the heavy dust load from the spray dryer. The on-line pulse cleaning system cycles less than once per hour to maintain the pressure drop setpoint.

Few problems associated with this fabric filter installation have been noted. Minor corrosion has been noted around doors and inspection covers. The hoppers have remained clean, with no dust build-up over time. Minor corrosion has also been noted near the hopper outlets. The Nomex filter bags have achieved more than three (3) years of life. Corrosion at one time caused the failure of several bag retaining rings, allowing those bags to drop off the bag cages. Improved bag hardware has corrected this problem.

7. SYSAV

The fabric filter installation at the SYSAV incinerator continues to operate on a year-round basis without problems. The inlet dust load, even after the dry reactor, is quite low at this plant, 0.5 gr/SCF, due to the action of the settling chamber. Filter pressure drop is maintained at five (5) inches W.G. with a pulse cleaning frequency of 2-4 cycles per hour. Only one incidence of hopper dust build-up and pluggage has been reported since 1981.

The only baghouse modification required for incinerator service was due to chlorine corrosion of the pulse valves. These valves are cold relative to the flue gas and can condense HCl. The modification was successful and no other problems have been reported.

IV. FABRIC TESTING

Fabric testing data is available from several of the baghouse installations. Used filter bags were removed after various service periods and tested for strength and filtration properties. This data is presented in Table 4. In general, all bags from dry scrubber installations indicate low to medium dust retention, low nodula formation and acceptable permeability after several years of service.

The bag removed from the University of Minnesota baghouse had the longest length of service (52 months), yet had lost less tensile strength than any other low ratio bag tested. It also had the highest permeability. Examination of the fabric and dust cake revealed very low levels of nodula formation. The dust cake was found to be very porous and had excellent release properties. Soxhlet extraction of nodulas indicated that the base component was calcium sulphate. The alkaline pH value demonstrates that no chemical degradation is occurring. Continued operation of this baghouse at the current conditions should result in very long bag life.

The bag removed from the Rockwell International installation had been in service only seven (7) months. The top three (3) feet of the bag had a high nodula population. The remaining bag had a low nodula population and low retained dust. Permeability in the top section was only 2.31 CFM compared to 3.75 below this zone. Some loss of breaking strength was found in this zone also. The nodula formation and loss of strength was traced to the chemistry of the deposits. Deposits on this bag were found to have a pH of 3.46. This acidic condition was unexpected after a scrubber. Iron sulphate and aluminum sulphate were predominant in the deposits, probably due to the reaction of SO_2/SO_3 with the fly ash. This acidic conditon was the cause of the measured decrease in bag strength. It is theorized that since the Rockwell Installation typically operates at 300 to 400 ppm SO_2 after the baghouse, excess alkalinity may not exist at all times in the ash handled. High moisture, low temperature conditions wihtout excess alkalinity present may exist during frequent boiler warm-ups. Acidic nodulas may form at this time and prevent the alkaline ash from neutralizing the acid during normal operations. A slightly higher lime feed rate to the scrubber should correct this problem.

The bags removed from the North Valmy Plant provide some comparison of scrubbed and non-scrubbed flue gas service. However, due to the much longer length of service with Unit #1, direct comparisons should be made with caution. The bag removed from Unit #1 had lower permeability and much higher nodula formation than the bag from Unit #2. Surface examination and Soxhlet extractions of both bags indicated basic differences in the bag's environment. Unit #1 fabric had many fibers totally encapsulated with calcium and aluminum sulfate salts. Unit #2 fabric had much less fiber encapsulation and the salts present were highly enriched with calcium sulfate. This type of nodula provides better fabric protection and promotes easier dust release. This would indicate that the scurbber is protecting the baghouse to a considerable extent and that longer fabric life can be expected. (3)

One bag was tested from the Ogden/Bristol installation to establish if any degradation had occurred during start-up. No evidence of any problems were found.

Felted fabric testing has also been done at the SYSAV installation in Sweden. Test bags with fifteen (15) combinations of batt and scrim fiber (Nomex, Ryton, PTFE and glass) were placed into service in this baghouse. Over a two (2) year period, bags were periodically removed and tested. Bag failures during the test period were recorded. Results indicated that only Nomex/PTFE (batt/scrim) and Ryton/PTFE had life-times exceeding the two (2) year test period. Indications are that use of PTFE scrim will extend bag life by 2.0 to 2.5 times. Evaluation of various testing methods also found that the acid cycle test gives no indication of relative baglife. Reference No. 3 contains a detailed report of this bag testing.

V. CONCLUSIONS

Fabric filter installations handling flue gas from wet/dry and dry scrubbers have a unique operating environment that presents special problems and opportunities. The following conclusions can be drawn from the baghouse installations surveyed.

1. Process conditions require operation at lower temperatures and higher humidity than normal ash collection-only duty. There will always be the danger of fabric blinding from wetting, although this is not a common problem. However, operating experience has shown that with the proper process controls and operator attention, fabric filters can survive and provide excellent dust control as part of scrubber installations.

2. Operation at low temperature and high humidity requires special attention in design and maintenance to avoid corrosion and dust deposits from condensation in cold areas. Designs utilizing the best practical insulation and sealing systems are mandatory. Adequate hopper heating is required to prevent dust handling difficulties. Special maintenance attention must be given to all sealing points. Protective coatings, when properly specified and applied, have greatly reduced this problem.

3. Filtering and handling large quantities of light, porous, hydroscopic dust requires special attention. The ash evacuation system must be sized to handle large surges of dust. Ash must be moved quickly and should not be stored in the baghouse hoppers. The bag cleaning system must be flexible enough to respond to changes in dust quantity and characteristics.

4. Woven fiberglass filter bags coated with Teflon provide excellent performance and long life in handling treated flue gas. Scrubbers generate large volumes of alkaline dust which tends to protect the fiberglass bags from chemical attack. The low cleaning frequency at most installations tends to reduce mechanical wear on the filter fabric, leading to very long service life. Fabric filters have an excellent service record in controlling emissions and maintaining low opacity in dry dry scrubbing applications.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

ACKNOWLEDGEMENTS

The authors wish to acknowledge the contributions and co-operation of administrative, operational and maintenance personnel at all of the baghouse installations surveyed.

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3. Environmental Consultant Company Report TLM 5935, Furnished by Sierra Pacific Power Company and Idaho Power Company, December 21, 1987.

TABLE 1
SURVEY INFORMATION

PLANT NAME	LOCATION	SERVICE	NO. UNITS, TYPE	FUEL SOURCE % S	EQUIP	FABRIC	CLEANING METHOD	START UP
1. Univ. of MN	Minneapolis Minnesota	CFB	1, Stoker 1, PC	Wyoming 0.5	LRFF	Glass	RA	09/83 07/84
2. Rockwell Internat'l	Columbus, Ohio	CFB	4, PC	Ohio 1.0 - 3.5	LRFF	Glass	RA	12/85
3. North Valmy	North Valmy, Nevada	CFB	2, PC (*)	Utah 0.4	LRFF	Glass	RA	1980 04/85
4. Ogden/ Bristol	Bristol, Connecticut	RRF	2, Stoker	Municipal Waste	HRFF	Glass	Pulse, Off-Line	10/87
5. Quebec City	Quebec City, Quebec	RRF	Stoker	Municipal Waste	HRFF	Glass Teflon	Pulse, On/Off Line	08/86
6. City of Sodertalje	Sodertalje, Sweden	CFB	3, PC	Various 0.7 - 1.5	HRFF	Nomex	Pulse, On-Line	10/82
7. SYSAV	Malmo, Sweden	RRF	Stoker	Municipal Waste	HRFF	Nomex, Ryton, Glass	Pulse, On-Line	1981

NOTES: (*) #1 Boiler Baghouse Only, #2 Boiler With Dry Scrubber

ABBREVIATIONS:
CFB = Coal Fired Boiler
PC = Pulverized Coal
RA = Reverse Air
RRF = Resource Recovery Facility
LRFF = Low Ratio Fabric Filter
HRFF = High Ratio Fabric Filter

TABLE 2
BAGHOUSE DESIGN DATA

PLANT NAME/ EQUIPMENT	O/A DIMENSIONS L x W x H (FT)	BAG DIA	NO. OF BAGS PER COMPARTMENT	NUMBER OF COMPT'S	GAS FLOW ACFM @ DEGREES F	FILTER SPEED FT/MIN NET (*)
1. University of Minnesota/ LRFF	52 x 36 x 46	8"	133	8	97,000 @ 150	2.5
2. Rockwell International/ LRFF	54 x 50 x 47	8"	149	8	105,000 @ 170	2.1
3. N. Valmy/ LRFF	140 x 104 x 81	12"	648 (#1) 612 (#2)	10 10	1,246,000 970,000	2.3 2.0
4. Ogden - Bristol/HRFF	27 x 30 x 41	6"	266	4	65,500	3.7
5. Quebec City/ HRFF	9′ x 5′ x 25′	5"	72	1	4,500	4.9
6. City of Sodertalje/ HRFF	41′ x 25′ x 66′	5"	540	4	129,300	4.6
7. SYSAV/HRFF	N.A.	5"	308	4	127,000	4.8

NOTES AND ABBREVIATIONS:

(*) NET DESIGN:
1 Compartment Off Line - LRFF
1 Compartment Off Line - HRFF, Off-Line Cleaning
0 Compartment Off Line - HRFF, On-Line Cleaning

N.A. = Not Available

TABLE 3
CURRENT TYPICAL BAGHOUSE OPERATING CONDITIONS

PLANT NAME/ EQUIP	INLET FLOW ACFM	INLET TEMP F	TYPICAL TEMP LOSS, F	ESTIMATED INLT DUST GR/ACF	FILTER SPEED, FPM/NET	AVG. PRESS LOSS IN W.G.	CLEANING INTERVAL, HOURS	STACK OPACITY %	MEASURED OTLET DUST GR/ACF
1. Univ. of MN/ LRFF	Unit #3 62,000	195	5	2.0	1.7	3.0	15	4	0.004
	Unit #4 96,000	220	5	1.0	2.5	2.9	5	3	0.008
2. Rockwell Int'l/LRFF	97,000	195	15	2.0	2.0	3.5	30	3	0.006
3. North Valmy/ LRFF	Unit #1 933,000	282	2	2.5	1.8	6.5	Cont	5	<0.006
	Unit #2 868,000	180	2	9.0	1.8	5.8	0.5	2	N.A.
4. Ogden-Bristol/ HRFF	70,000	285	10	2.7	4.1	6.0	Cont.	2	N.A.
5. Quebec City/HRFF	4,100	285	10	2.5	4.7	5.9	Variable	N.A.	<0.006
6. City of Sodertalje/ HRFF	131,000	160	2	5.0	4.7	4.8	1	N.A.	<0.006
7. SYSAV/ HRFF	127,000	300	N.A.	0.4	4.9	5.0	0.4	N.A.	<0.006

NOTES & ABBREVIATIONS:
N.A. = Not Available
< = Less Than

TABLE 4

FABRIC TESTING DATA

PLANT NAME/ EQUIPMENT	FABRIC/ FINISH	LENGTH OF SERVICE, MONTHS	PERMEABILITY CFM/SQ.FT @ 0.5" W.G.		BREAKING STRENGTH, % LOSS		FLEXCYCLES, % LOSS		RETAINED DUST, H:HEAVY, M:MEDIUM, L:LIGHT	PH
			AS REC'D	VACUUMED	LGTH	WDTH	LGTH	WDTH		
1.Univ. of of Minn./ LRFF	Glass/ Teflon	52	5.00	15.77	2.26	6.13	18.34	50.15	L	8.15
2.Rockwell Int'l/ LRFF	Glass/ Teflon	7	3.43	13.47	13.86	16.19	40.55	47.73	L	3.46
3.North Valmy/ LRFF										
Unit #1	Glass/ Teflon	84	2.00	5.56	45.90	41.56	55.15	85.46	H	10.30
Unit #2	Glass/ Teflon	34	3.40	8.93	11.14	22.44	19.87	38.16	M	11.12
4.Ogden/ Bristol/ HRFF	Glass/ Teflon	2	7.93	22.94	0.0	0.0	2.39	15.29	L	6.17

DUST COLLECTOR DESIGN CONSIDERATIONS FOR MSW ACID GAS CLEANING SYSTEMS

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ABSTRACT

Spray Dryer Absorption (SDA), since first applied to coal-fired utility boilers in 1979, has gained rapid acceptance as a cost effective, reliable process for removing SO_2 and other acid gases from boiler flue gas. More recently the increased use of incinerators, to simultaneously reduce the volume of municipal solid waste (MSW) and recover energy, has lead to applications of SDA for the simultaneous control of HCl, HF, SO_2, other acid gases, trace compounds, and dust emissions from incinerator flue gases.

In utility SDA flue gas desulfurization applications, flue gases exiting the spray dryer have substantially different gas conditions as compared to the inlet. The same holds true for MSW applications, however, as compared to utility applications, flue gas temperatures are typically higher, grain loadings lower, and the SO_2 content is lower. Some HCl is present and the nature of the dust is quite different. These changes in flue gas conditions require a re-examination of both fabric filter and electrostatic precipitator design considerations to ensure successful applications.

This paper presents a discussion of fabric filter and electrostatic precipitator design considerations for SDA flue gas cleaning applications. Impacts of SDA conditioned flue gas on design requirements are reviewed. Pilot and full-scale test results and operating data for fabric filter and ESP applications are presented.

INTRODUCTION

Incineration has long been employed as a method to reduce the volume of municipal solid waste (MSW) which is disposed of in land fills. Traditionally these incinerators have been equipped with electrostatic precipitators as their air pollution control device. More recently concern over emissions of other pollutants, such as acid gases (HCl, SO_2, HF), trace organic compounds (dioxins, furans), and trace metals (lead, cadmium, mercury), has lead to the promulgation of regulations to control these emissions. Spray Dryer Absorption (SDA) has been widely specified and selected as best available control technology for meeting emission control requirements.

In SDA, an alkaline slurry is sprayed into the flue gas to simultaneously cool the gas and react with acid gases present to form solid particulate matter. A dust collector is located downstream of the alkali injection point to remove the particulate matter and allow for additional acid gas removal. The alkali slurry is typically introduced into the flue gas in a spray dryer and the dust collector is either a fabric filter or an electrostatic precipitator (ESP).

Flue gas conditions after the spray dryer differ significantly from those leaving the boiler. The flue gas temperature is lowered, the moisture content is increased, acid gas concentrations reduced, the particulate loading is increased, and the nature of the particulate is altered. These changes in flue gas conditions require a re-examination of both fabric filter and electrostatic precipitator design considerations to ensure successful operation of either dust collection device.

This paper presents a discussion of fabric filter and electrostatic precipitator design considerations for SDA acid gas cleaning applications. The SDA acid gas cleaning process is described. Changes in flue gas conditions are reviewed and recommended sizing criteria are presented. Results from pilot plant and full-scale emission tests are given.

SYSTEM DESCRIPTION

The SDA acid gas cleaning system generally consists of a spray dryer absorber and a dust collector per MSW incinerator with a common reagent preparation system for multiple incinerators. Figure 1 shows a simplified flow diagram for the process.

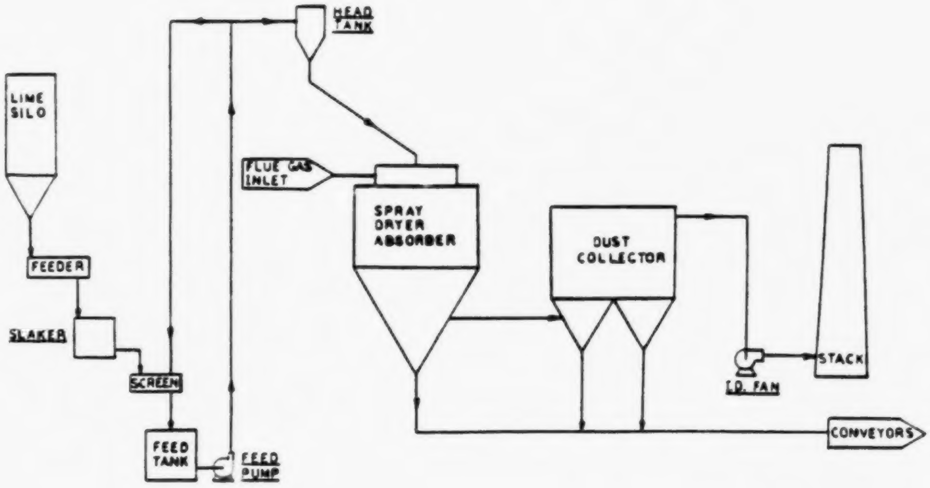

Figure 1. Simplified Flow Diagram

For larger installations pebble quick lime (CaO) is usually used as the alkaline reagent. Quick lime is delivered by truck or rail, and conveyed pneumatically to a storage silo. From the silo lime is metered into a slaker where enough water is added to produce a 20-25 percent hydrated lime ($Ca(OH)_2$) slurry which is screened to remove grit, and then flows into a lime storage (or feed) tank. From the lime tank the slurry is pumped to a head tank where the major portion overflows back to the lime tank via a screen. Dilution water is added at the lime tank, or head tank, to produce a slurry with the proper alkalinity to achieve the required acid gas (HCl, SO_2, SO_3, HF) removal. The dilution can be manually or automatically controlled based on a feedback signal from continuous monitors in the stack.

For smaller applications hydrated lime is often used as the reagent, thus eliminating the requirement for a lime slaker.

A portion of the diluted lime slurry flows via a temperature control valve from the head tank to the rotary atomizer where it is converted into a cloud of fine droplets with a very high surface area for drying and acid gas reactions. The spray dryer outlet temperature, for a given inlet flue gas flowrate and temperature, is directly proportional to the amount of water entering the atomizer and can be controlled within a few degrees fahrenheit of the set point.

The lime slurry enters the spray dryer via the atomizer where it mixes with the hot incoming flue gas, simultaneously reacts with acid gases present and dries prior to exiting the dryer. The majority of reaction products formed are carried over to the dust collector where additional acid gas removal takes place. Dry product collected in the SDA and dust collector is then conveyed mechanically or pneumatically to a waste product silo, or can be conveyed to the bottom ash quench tank for mixing with the bottom ash. A portion of the dry product can be recycled back to the feed system to improve reagent consumption and acid gas removal if economically justified.

For efficient acid gas absorption and lime utilization, the spray dryer outlet temperature is controlled at as low a temperature as possible but high enough to ensure a dry, free-flowing product under all operating conditions.

The dust collector downstream of the spray dryer may be an electrostatic precipitator, a reverse-air baghouse or a pulse-jet type baghouse. The selection of a specific type of dust collector is dependent on site specific factors such as particulate emission limits, overall acid gas removal requirements, project economics and client preference. Each of these dust collection devices offers process advantages and disadvantages that are evaluated on a site specific basis.

The flue gas composition changes significantly as it passes through the spray dryer. Table 1 illustrates differences the dust collector sees with and without a spray dryer upstream.

TABLE 1. TYPICAL FLUE GAS CONDITIONS

	Entering Spray Dryer	Leaving Spray Dryer
Flue Gas Temperature, °F Dry Bulb	350 - 450	240 - 290
°F Wet Bulb	130 - 140	135 - 150
Flue Gas Composition:		
Moisture, % volume	8 - 16	12 - 20
HCl, ppm	400 - 1000	50 - 200
SO_2, ppm	150 - 500	20 - 250
HF, ppm	5 - 25	1 - 5
Particulate, gr/acf	0.5 - 6.5	3 - 9
Relative Gas Volume, acfm/acfm	1	0.85 - 0.95

As is shown, the dry bulb temperature decreases while the wet bulb temperature increases due to the increased moisture content, however, the approach to saturation remains 90-100°F above the dewpoint. Acid gas concentrations are reduced by 50-90 percent and particulate loadings are increased by about 1.5 to 3 grains/acf depending on lime stoichiometry and acid gas removal efficiency. The actual gas volume is decreased by about 5-15 percent, depending on inlet and outlet temperatures.

In addition to an increase in the particulate loading leaving the spray dryer, the nature of the particulate is changed due to the addition of lime and the reaction products formed from the acid gases removed. Table 2 presents ranges of selected components and properties for particulate matter entering and leaving the spray dryer.

TABLE 2. PARTICULATE MATTER PROPERTIES

	Entering Spray Dryer	Leaving Spray Dryer
Composition, % wt		
Ash Fraction	80 - 95	50 - 80
$Ca(OH)_2$	0 - 15	5 - 25
$CaCl_2.x\ H_2O$	0 - 5	5 - 15
$CaSO_x$	0 - 10	5 - 10
H_2O, (total)	< 5	5 - 20
Loss on Ignition @ 900°C, %	10 - 30	15 - 40
Particle size, 50% less than μ	25 - 45	15 - 30
Resistivity, ohm-cm	$10^9 - 10^{12}$ @ 300 - 400°F	$10^8 - 10^9$ @ 290°F

The material entering the spray dryer is primarily fly ash carried over from the incinerator whose composition varies substantially with the nature of the refuse being fired. Some hydrogen chloride in the gas stream can react with alkaline material in the ash to form sodium and calcium chlorides, producing a sticky ash. The ash generally has a high loss on ignition due to carbon carryover and the presence of thermally unstable compounds in the ash. The average particle size distribution can vary widely, depending on the refuse fired and the incinerator design. The resistivity of the ash typically shows a peak value between 300 and 400°F. Resistivities can range from less than 10^9 to greater than 10^{12} ohm-cm at 400°F.

The product after the spray dryer has a lower ash fraction and higher fractions of lime, calcium chloride, calcium sulfites and sulfates, and moisture. The average particle size is lowered due to the addition of very fine lime particles. The product tends to be stickier than the fly ash. Due to the increased fraction of calcium chlorides present, the product has a hygroscopic nature. The total moisture content (free plus chemically bound) can range from 5-20 percent. The resistivity is typically about 10^9 ohm-cm in the 240-290°F range.

These changes in flue gas conditions and nature of the product after the spray dryer require special design considerations to ensure satisfactory long term operations of the dust collector. Factors which must be addressed include corrosion, product stickiness, hydroscopicity, and the increased amount of product.

GENERAL DESIGN CONSIDERATIONS

Whether a fabric filter or ESP is selected as the dust collector, minimization of heat loss from the dust collector to avoid corrosion and increased product stickiness is a prime design consideration. Four methods employed to achieve this are:

- Insulation
- Control of Air Inleakage
- Hopper Heating
- Operating Procedures

Because the dust collector sees flue gas temperatures much closer to the dew point temperature, rugged insulation specifications are needed to both minimize overall temperature drop and low local temperatures that may contribute to localized corrosion. Insulation thickness should be a minimum of four inches with double lapping on housing side panels. Hopper crotch areas should be insulated and hopper support columns isolated or insulated to minimize "cold finger" effects. Very low outlet temperatures (230-240°F) should include insulation between compartments of fabric filters.

Air inleakage can contribute to overall temperature drop and localized cold spots. Good quality control during fabrication and erection of the dust collector is essential to ensure good fit of components and full seal welds of all joints. We recommend the filters be pressureized and leak tested using smoke bombs at the completion of erection. The number of access doors should be minimized to reduce potential air inleakage points. Special access door designs and gasketing materials should be used to further minimize potential inleakage.

Hopper heating is necessary to minimize heat loss and to maintain flowability of the product. Low watt density heaters should be applied to as much area of the hopper as is economically reasonable. The heaters should be sized to maintain skin temperatures the same as product or gas temperature. For most applications, power densities of 300-400 watts per square foot applied to the bottom one-third of the hopper are adequate.

Operating procedures for dust collectors, particularly fabric filters, should address removal of the product and purging on system shut down. The sticky nature of the product can increase with time, therefore, when a fabric filter is shut down the compartments should continue to go through at least one complete cleaning cycle. For ESP rapping should be continued long enough to remove collected product. Because of the high moisture content and acid gases in the flue gas, it is necessary to adequately purge the dust collector when shuting down. If a fabric filter compartment is removed from service for other than cleaning, it should be purged by opening access doors and then closed back up to minimize heat loss across internal walls.

When the incinerator is started-up, the fabric filter should be bypassed, or the ESP should be left de-energized until the flue gas temperature is high enough to allow the spray dryer to be put into service. This allows for a protective coating of the internals by the dry alkaline product as acid gases are introduced.

The ash handling system for the dust collector must be sized to handle higher dust loads and different product transport properties than for normal ash applications. Continuous product evacuation from the hoppers is recommended. Mechanical conveyors, particularly en mass type, have demonstrated high reliability in this service. If pneumatic conveying systems are used, consideration should be given to heating the transport air to approximately the operating flue gas temperature to minimize potential transport system plugging problems. Insulation of pneumatic transport lines and filter receivers, or indoor installation should be considered in locations that experience low ambient temperatures.

FABRIC FILTER DESIGN CONSIDERATIONS

The primary design decision concerning fabric filters is whether to select a reverse air cleaned or a pulse cleaned baghouse. This selection is based on site specific economics and dust emission requirements.

The reverse air cleaned baghouse uses a low energy cleaning. In a reverse air baghouse the flue gas is introduced through a manifold into the hopper section of the various compartments. The flue gas then turns up across the tube sheet and into the filter bags. After passing through the filter bags, the flue gas exits through an automatic outlet valve into the outlet manifold and to the ID fans and the stack. The dust cake remains on the inside of bags.

Once the dust cake on the filter bags builds up to a point where the resistance to flow is unacceptable, the filter bags must be cleaned. In a reverse air collector this is done by first closing the outlet valve, then opening a reverse air valve and allowing air to flow in the opposite direction from the outside of the bags to the inside. This causes the bag to collapse in a star shaped pattern, creasing the fabric, fracturing the filter cake and dislodging the cake from the filter bags allowing the collected particulate to fall into the collection hoppers for disposal. Figure 2 shows JOY's Therm-O-Flex® reverse air cleaned fabric filter.

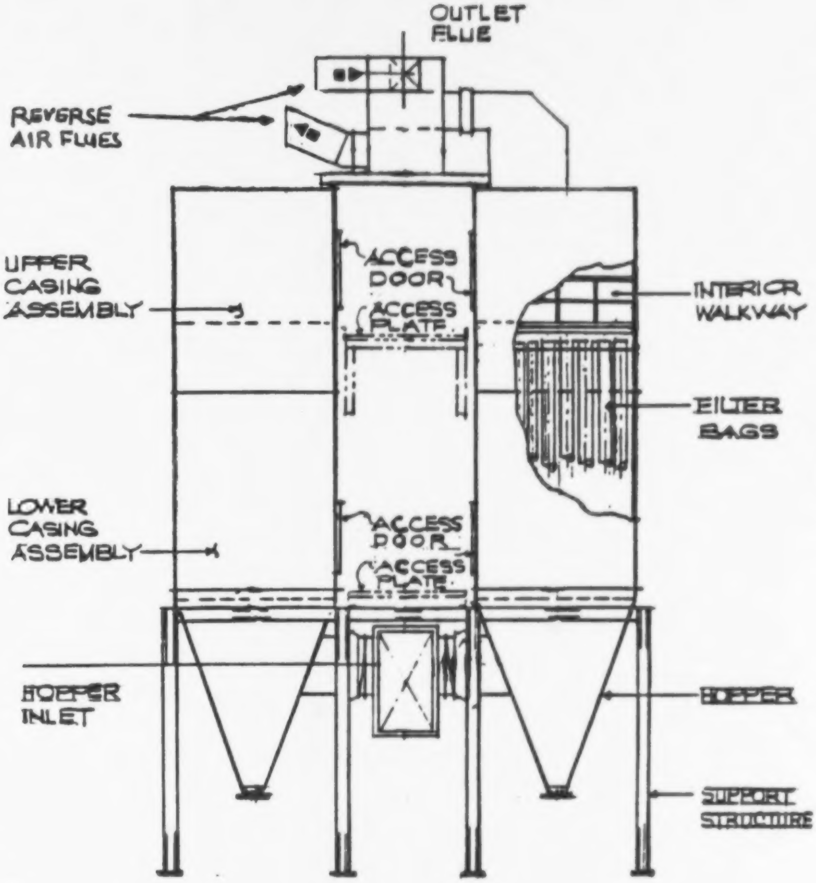

Figure 2. Therm-O-Flex Fabric Filter

In contrast to the reverse air baghouse, the pulse cleaned baghouse uses higher energy cleaning. The flue gas is once again distributed through a manifold into individual collector modules. However in the case of a pulse cleaned collector, the particulate is collected on the outside of the filter bags which are supported using metal cages. The flue gas passes through the filter media into the inside of the filter bag through the tube sheet, out the automatic outlet valve and into the outlet manifold. When resistance across the filter media requires the filter bags to be cleaned, a module is taken off-line with the closure of the outlet valve for that individual module. Bags are then cleaned one row at a time with the introduction of compressed air into the top of the filter bag causing the bag to flex away from the cage, dislodging the particulate on the outside of the bag, which then falls into the hoppers for disposal. Figure 3 shows JOY's pulse cleaned fabric filter.

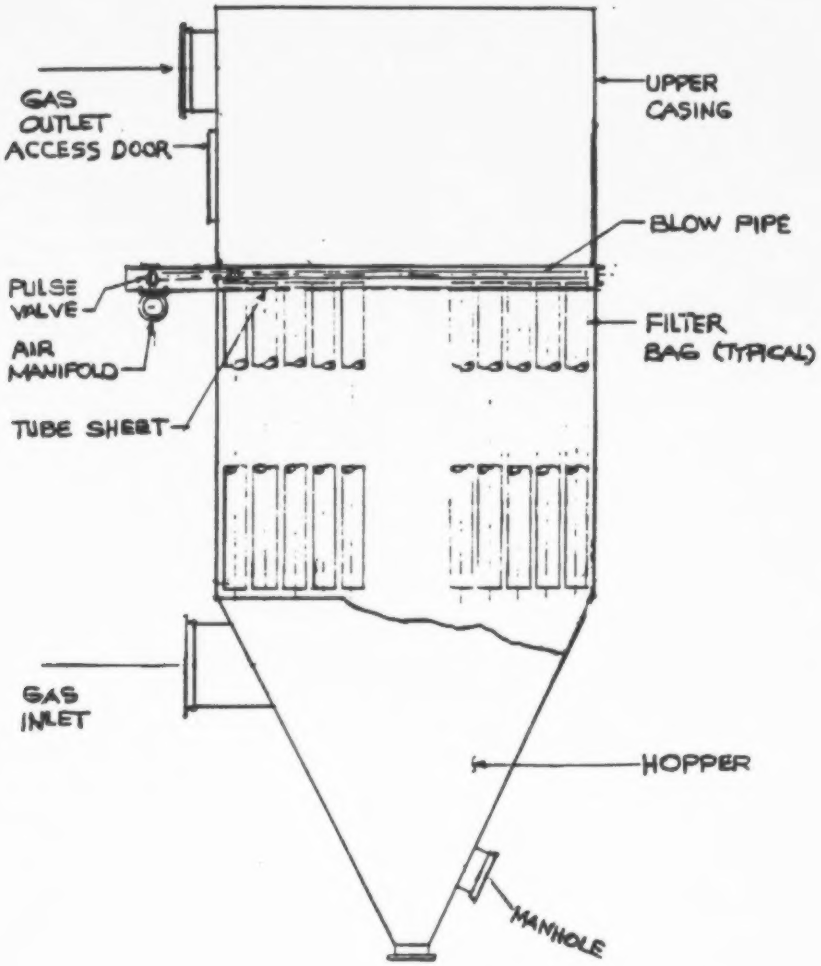

Figure 3. Pulse Cleaned Fabric Filter

The JOY pulse cleaned baghouse, called the PULSEFLO®, utilizes relatively low pressure air for cleaning. The low pressure cleaning system utilizes air in the 40-60 psi range versus the more prevalent high pressure system which utilizes air pressures in the 80-100 psi range. The lower air pressure allows for a more efficient cleaning system by utilizing larger pulse valves and graduated internal headers which allow for the use of substantially longer filter bags. The design incorporates a walk-in outlet plenum to minimize air inleakage as compared to a hatch cover design.

Table 3 presents recommended design parameters for fabric filters in SDA acid gas cleaning applications.

TABLE 3. FABRIC FILTER PARAMETERS.

	Reverse Air	PULSEFLO®
Bag Material	Woven Fiberglass	Woven Fiberglass
Fabric Coating	Minimum 10% Teflon B	Minimum 10% Teflon B
Fabric Wieght, oz/yd^2	9.5	16 or 22
Bag size, Ø" x L'	8 x 24	6 x 20
Maximum Air-to-Cloth Ratio, net	2:1	4:1
Estimated Bag Life, years	4	2
Minimum Compartments	6	6

Fiberglass bags with a Teflon B coating are recommended for these applications because of their excellent acid resistance, and their ability to operate continuously at up to 500°F. The fabric weight is heavier for the PULSEFLO® filter to give higher mechanical wear for the higher energy cleaning.

The major advantage of pulse cleaned baghouses is that they operate at relatively high filtering velocities in contrast to reverse air baghouses. The PULSEFLO® operates at a filter velocity of approximately four to one while the reverse air operates in the range of two to one. This results in a much smaller baghouse and corresponding lower first costs. Disadvantages of the pulse cleaned unit are that at the higher filtering velocity, emission levels and pressure drop can be expected to be slightly higher than in the reverse air unit.

Another disadvantage of the pulse cleaned unit is a shorter expected bag life compared to a reverse cleaned unit, resulting in higher life time maintenance costs. Savings in first costs must be evaluated against increased bag costs and pressure drop over the life of the system in order to determine the economic choice.

As stated above, emissions from pulse cleaned baghouses are generally higher than for reverse air cleaned units with the same type of bags. The exception to this is if Goretex® membrane filter bags are employed. Table 4 presents results of emission measurements made on pilot and full-scale SDA gas cleaning applications.

TABLE 4. FABRIC FILTER PARTICUCLATE EMISSIONS

Type		No.	EMISSIONS		
Filter	Fabric	Tests	Range, GR/ACF	AVE. GR/ACF	GR/SCFD @ 12% CO_2
Reverse Air	Woven Fiberglass	18	0.0010-0.0032	0.0019	0.0054
PULSEFLO®	Woven Fiberglass	6	0.0025-0.0065	0.0040	0.0114
PULSEFLO®	Gortex	6	0.0009-0.0027	0.0015	0.0043

As is shown, average emissions for the PULSEFLO® filter are approximately double those for the reverse air with woven fiberglass bags. The PULSEFLO® equipped with Gortex® bags has the lowest average emissions, however, it should be noted that Gortex® bags cost approximately five times that of fiberglass. Also shown is a comparison of the measured average emissions (GR/ACF) to typical regulatory values (GR/SCFD @ 12% CO_2).

A minimum of six compartments is recommended for either a reverse air or PULSEFLO® fabric filter in SDA acid gas cleaning applications. This is based on minimizing system pressure drop swings as compartments are taken off-line and cleaned. In addition, as cleaned compartments are put back into service, the acid gas removal efficiency (especially for SO_2) is initially low for that compartment. As the number of fabric filter compartments increases, the impacts of off-line cleaning on overall system performance are minimized.

ESP DESIGN CONSIDERATIONS

ESP design parameters which must be considered for SDA acid gas cleaning applications include:

- Sizing Criteria
- Electrode Systems
- Rapping Systems
- Insulator Compartment Ventilation

Sizing criteria for these applications are selected based on data collected from pilot and full-scale precipitators following spray dryers for acid gas cleaning applications. Data collected during parametric and performance tests have included: specific collecting area, gas velocity, secondary voltage and current, inlet and outlet dust loadings, dust particle size, resistivity over temperature and moisture range, ash compositions, and opacities. Using this data base and the modified Deutsche equation, sizing parameters are selected for each application. Table 5 presents typical ranges of sizing parameters used for U.S. acid gas cleaning applications.

TABLE 5. ELECTROSTATIC PRECIPITATOR - SIZING PARAMETERS

Required Efficiency, %	98 - 99.9
Number of Fields	3 - 4
SCA, ft^2/1000 acfm	400 - 500
Average Secondary Voltage, kv	35 - 55
Average Secondary Current, mA/1000 ft^2	30 - 50
Gas Velocity, ft/sec.	3.0 - 3.5

The data shown are typical for achieving emission levels of 0.01-0.015 GR/SCFD corrected to 12 percent CO_2 based on operation with the spray dryer in service. JOY's sizing model indicates that the same or slightly higher emission levels will occur with the spray dryer out of service due to the differences in ESP operating conditions: i.e., higher gas volume and velocity, lower gas moisture, and higher dust resistivity. Table 6 presents emission data from three European applications where SDA's are retrofitted upstream of existing ESP's.

TABLE 6. ELECTROSTATIC PRECIPITATOR - PARTICULATE EMISSIONS, GR/ACF

			MEASURED	
Site	Number Fields	Guarantee	Without Spray Dryer	With Spray Dryer
Zurich, Josefstrasse	2	0.01	0.0022	< 0.0022
AVG Hamburg	2	0.015	0.009	0.008
City of Leverkusen	3	0.01	< 0.0006	< 0.0004

As is shown, there is almost no difference in emissions with or without the spray dryer in service. With the spray dryer in service, the measured SCA (specific collecting area) at Zurich during the performance tests was approximately 425. It should be noted that in all cases the tested emissions are well below the guarantee values.

Electrode systems for acid gas cleaning applications must be designed for the corrosive gas conditions and the sticky nature of particulate being collected. JOY currently recommends a rigid frame electrode discharge system with spiral discharge electrodes fabricated of Incoloy 825. Figure 4 shows this design.

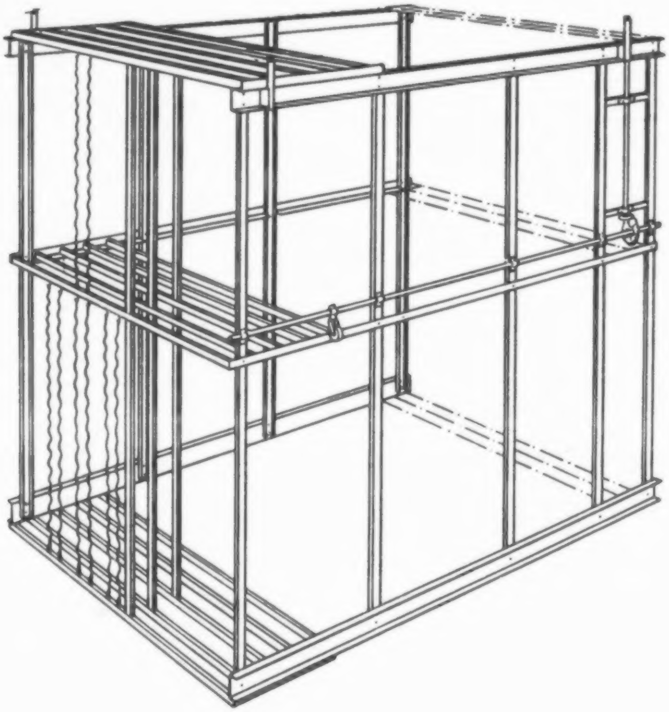

Figure 4. Rigid Frame Discharge Electrode System

This design is such that the rigid frame will not distort under normal operating conditions or short term temperature excursions, thus minimizing potential alignment problems.

For the collecting electrode system, a heavy gauge roll formed collecting module is recommended. Figure 5 shows this type of collecting electrode system.

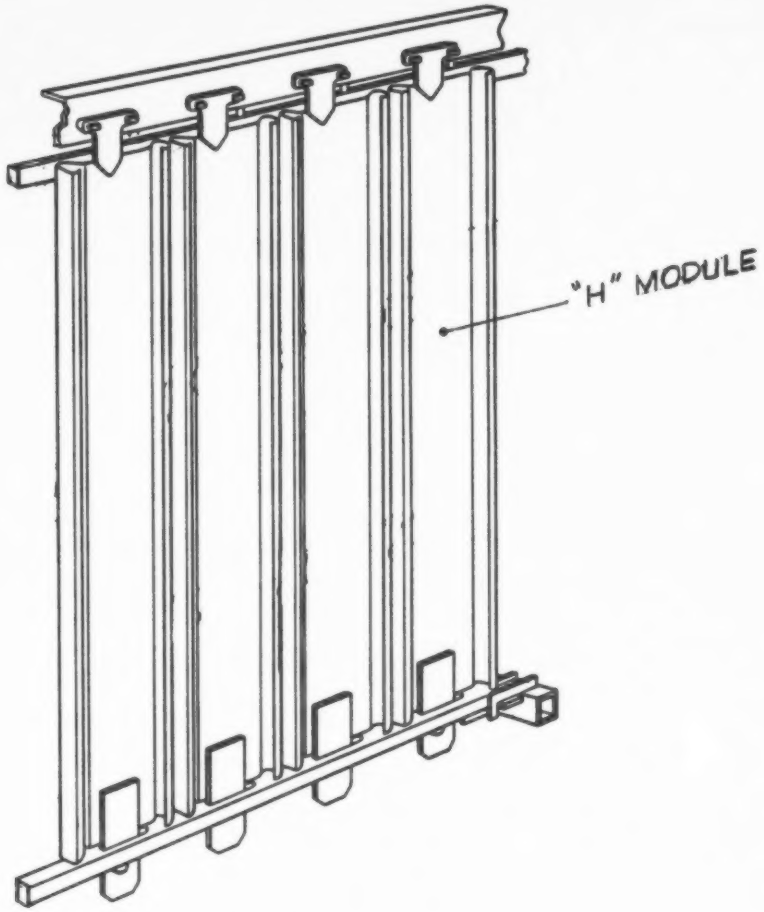

Figure 5. Collecting Electrode System

The collecting plates are eccentrically hung at the top and guided by a shock bar at the bottom. This allows individual plates to withstand normal or upset temperature excursions without warping. The design also allows for high rapping forces to be applied without distorting the collecting electrodes.

Tumbling hammer rappers are recommended for these applications where sticky dusts may be encountered. Figure 6 shows the rapping system for our rigid frame ESP.

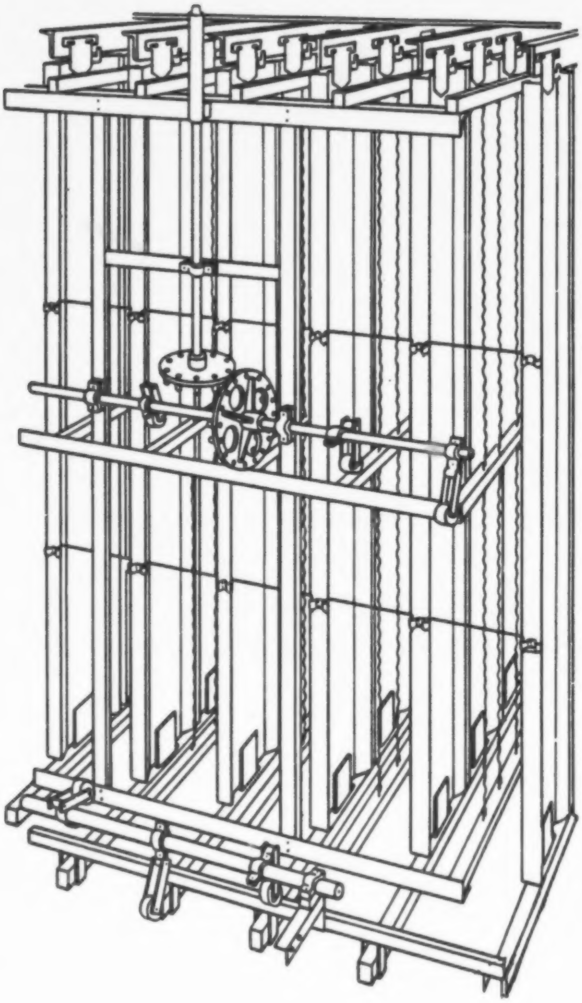

Figure 6. Rigid Frame Tumbling Hammer Rapping

The emitting rappers rap at the center of the frame, producing a rapping acceleration which is distributed evenly via a rapping angle to the individual spiral electrodes. This system minimizes misalignment problems. The collecting system is rapped from the bottom by tumbling hammers and each collecting plate is rapped by an individual hammer thus minimizing rapper losses.

The design of the insulator compartment ventilation system must consider the lowered flue gas temperature and increased moisture content. A pressurized ventilation system employing heated air is used to purge precipitator support insulator. This system is required to maintain clean insulators and reduce potential electrical tracking problems.

CONCLUSIONS

Both fabric filters and precipitators can be applied successfully to MSW acid gas cleaning systems if properly designed for the flue gas and particulate conditions after a spray dryer absorber.

In the case of the fabric filter design considerations include proper air-to-cloth ratio, attention to insulation and heat losses, minimizing air inleakage through proper design and construction techniques, the use of low watt density hopper heaters and proper operating procedures.

On the other hand, when sizing precipitators, the design must balance the effects of decreased gas volume and increased moisture content of the gases versus increased dust loading and compositional changes in the particulate. The electrode systems must be of a proven heavy duty design and a tumbling hammer rapping system should be used for sticky, high chloride dust. Proper heating and venting of insulator compartments will ensure long term reliable operation.

Both types of collector, if properly designed, can be successfully used for acid gas cleaning applications. The proper choice should be based on site specific economic and permitting factors.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

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A Study of a Reverse-Gas Fabric Filter Controlling Particulate Emissions from a 50 MW Atmospheric Fluid Bed Combustor

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ABSTRACT

To demonstrate the technical reliability and economic advantages of Atmospheric Fluid Bed Combustion (AFBC) when burning several different qualities of coal and operating the turbine system at an ultra super critical temperature (593°C), an extensive testing program was established for the 50MW AFBC Unit located at the Wakamatsu Thermal Power Station of the Electric Power Development Company, Ltd in Japan. This testing will be completed over a five year period and will be the basis for scaling up to a 350MW Commercial Plant. Portions of this testing include parametric testing of a Reverse Gas Cleaning Fabric Filter installed for control of particulate emissions from the 50MW AFBC demonstration unit. This paper addresses this Fabric Filter's design, start-up, performance parameters measured to date, and the plan for future testing. Data such as flue gas flow rates, inlet/outlet and particulate concentrations, pressure losses, particle sizes, fabric properties, etc. are described. Overall Fabric Filter durabilities i.e. filter bags, dampers, inside wall surfaces, etc. are also addressed.

INTRODUCTION

Over the past few years there has been an increased interest and development in the area of Fluid Bed Combustors particularly in plants sized for 150MW and smaller. This has led to the development of many different combustor designs. In Japan the Electric Power Development Company (EPDC) has installed a 50MW Babcock Hitachi Bubbling Bed Fluid Bed Combustor as a demonstration unit for this technology and for the basis of scaling up to a future 350MW Commercial Plant. The plant site for this project is the Wakamatsu Coal Utilization Technology Research Center (Wakamatsu Thermal Power Station) of EPDC, Ltd. The demonstration/test period is to span five (5) years from April, 1987 through March 1992.

The AFBC Plant is as shown in Figure 1. A combination of In-Furnace SOx removal, Flue Gas Denitrification by SCR, and particulate control by a fabric filter are some of the main features of this demonstration plant. This paper reports the test data and overall performance of this fabric filter. This fabric filter was designed and supplied by Gadelius KK/Flakt of Kobe, Japan.

Why a Fabric Filter?

The main reasons EPDC opted for the fabric filter technology was:

A) Particulate outlet emissions are not as highly sensitive to the properties of the ash as are electrostatic precipitators.

B) Particulate outlet emissions are not as highly sensitive to the quantity of the ash as are electrostatic precipitators.

C) High removal efficiency of sub-micron particulates.

D) Ability to improve removal efficiency by upgrading filter media.

E) Reverse air cleaning was selected because of the good operating performances experienced in the USA utility industry with similar equipment.

Fabric Filter Design

The fabric filter is a low ratio type employing woven glass filter bags and a reverse air cleaning system. It is a common wall panel design much like the fabric filters now controlling particulate emissions on the U.S.A. utility boilers. Table 1 describes the main specification of this filter, Figure 2 shows the schematic plan layout of the filter, and Figure 4 is photographs of the actual installation.

Filter Bags

The main complement of filter bags are woven glass bags similar to the bags normally used in fabric filters for USA utility boilers. In compartment 3A twelve (12) Gore-Tex bags and twelve (12) Micro-Tex bags were installed as comparison test bags. Table 2 describes the specifications of the original and these membrane bags.

Hoppers and Ash Handling System

Hoppers are designed with a 63 degree valley angle and are equipped with ash air blowing nozzles and electric heaters. Ash is evacuated from the hoppers by a dilute phase negative pressure pneumatic conveying system. To date, there have been no problems with evacuating the fabric filter hoppers.

Schedule of Events

The fabric filter has treated the flue gas continuously from January, 1987 including all start-ups and shut downs. Overall schedule of events are and will be as follows:

Completed Construction	10/86
Initial Start-up	10/86
Performance Test #1	02/87
Performance Test #2	11/87
Bag Test #1	05/87
Bag Test #2	09/87
Phase #1 Test Program	04/87 to 03/89
Phase #2 Test Program	04/89 to 03/92

Fuel Types

Through 1987 three different coals have been fired in the AFBC. Table 3 is the analysis of these different coals. Table 4 describes the ash collected by the fabric filter.

Fabric Filter Start-up

Just prior to initial start-up the filter bags were precoated by external injection of fly ash. In addition the bags were also substantially precoated with limestone powder generated during the re-mounting of the bed material in the fluidized bed. (The Fabric Filter was not bypassed during initial or subsequent start-ups, - it is always on-line including shut-down periods). Initial light-off was accomplished with light oil. Within approximately ten (10) hours a steady full boiler load was reached. There have been few problems with this fabric filter since start-up. Corrosion, high emission levels, gas leakages, or ash pluggages have been non-existent at this plant.

Fabric Filter Test Program

Test data during this plant demonstration is developed for the fabric filter as well as the other components of this AFBC plant. This data will be taken over a five year period from 04/87 through 03/92. The fabric filter test program is divided into phase 1 and phase 2 test programs. Table 5 details the outline of these programs.

Test Results and Operating Data to Date

Fabric Filter Performance

Table 6 Summarizes fabric filter performance/test data taken on February 23, 1987 and again on November 27, 1987. Outlet emissions were taken via the in-stack method similar to EPA method 17. Samples of the ash were extracted at the inlet of the fabric filter for determination of inlet ash concentration and particle size. Type of fuel and baghouse pressure losses are also summarized. It should be noted the outlet emissions tested are extremely low. Testing at such low levels should not be considered to be the norm for fabric filter installations or design requirements for future installations.

Bags

In May of 1987, two (2) woven glass bags were removed from the fabric filter and tested. In September of 1987 six (6) other bags were tested including two (2) Gore-Tex bags. Table 7 lists the results of the woven glass bags. Figure 6, microscopic photos, show the clean surface, dirty surface,and cross-section of the bags from the May test. Figure 7, microscopic photos, shows ash on the warp and fill yarns of woven glass during the September bag test. Figure 8, microscopic photos, compare the ash build-up on the woven glass bags and the Gore-Tex bags after one month of operation. During all of this testing chemical or physical bag deterioration was not observed. As testing proceeds and if the membrane bags indicate superior filtration properties one entire compartment will be rebagged with membrane bags.

Discussion

Particle Morphology and Size

Shown in Figure 5 is the shape of the particles collected in the fabric filter hoppers. From this AFBC plant the particles are more irregular shaped and pointed when compared against ash generated from a typical Pulverized Coal Fired Unit. This compares similarly with data reported by Southern Research Institute (1). Particle size was measured at 2.5u MMD when burning coal number 1 and 16u MMD when burning coal number 3 (Table 6). This does indicate particle size can vary with the type of fuel fired in an AFBC.

Pressure Loss

Figure 9 shows the average operation of the reverse air cleaning system and the resulting pressure losses across the flange to flange fabric filter, versus time. When the pressure losses through the plenums and dampers are subtracted from these losses a grid sheet pressure loss is determined. The grid sheet pressure losses during the test run in February, 1987 were from 1.0 inch w.g. (clean bags) to 3.4 inches w.g. (dirty bags) when all compartments were on-line. Spikes to 4.6 inches w.g. were experienced during cleaning. This calculates to a clean filter drag of 0.7 inches w.g./FPM to a high of 2.5 inches w.g./FPM when bags are dirty. During the testing in November of 1987 pressure losses across the grid sheet ran from 1.8 inches w.g. (clean bags) to 3.4 inches w.g. (dirty bags), all compartments on-line, with spikes to 4.7 inches w.g. during cleaning. This calculates to clean filter drag of 1.3 inches w.g./FPM to a high of 2.6 inches w.g./FPM when bags are dirty. Drag data developed at the AFBC demonstration plant compares favorably with reverse air units presently operating at the USA P.C. fired utility boilers (1). This data does however show a small increase in pressure loss was maintained across the cleaned bags from February through November. Whether this increase is temporary and caused due to normal bag seasoning or changes in fuel or other parameters is not determinable at this time. In any case these values are more than acceptable for proper fabric filter performance.

Conclusions and Summary

- For over one year this fabric filter has controlled particulate emissions to extremely low levels while operating at low pressure drop levels.
- There has been no physical or chemical deterioration of the filter bags or the fabric filter components during this time frame.
- Further testing will be conducted to determine long range effects and/or changes in performance of the fabric filter as parameters of the AFBC are varied.
- Additional feasibility studies will be made to determine potential reductions in size of a fabric filter when using membrane bags.
- Reverse air low ratio fabric filter technology performs admirably when applied to this AFBC process.

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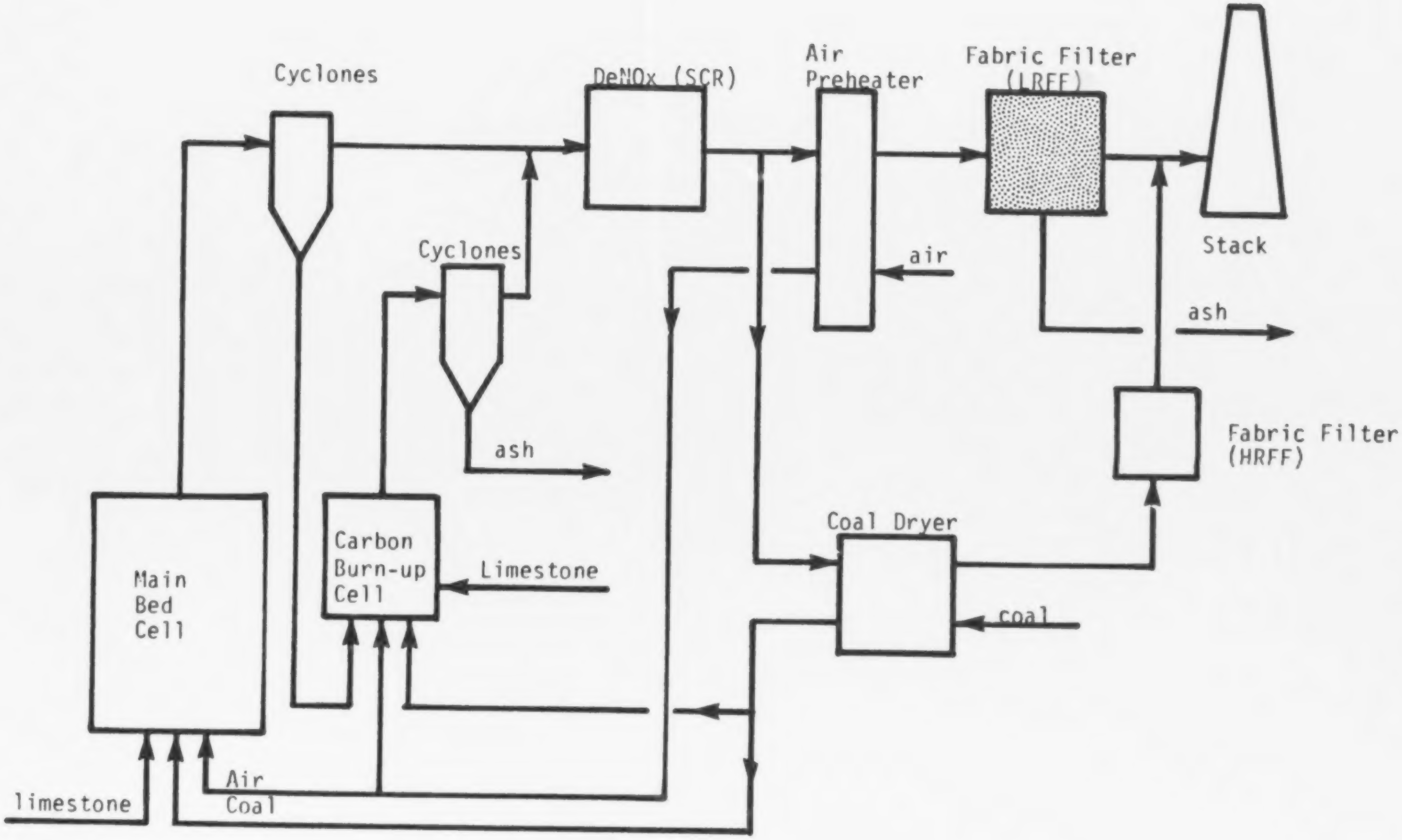


Fig.1 Schematic Diagram of AFBC Demonstration Plant

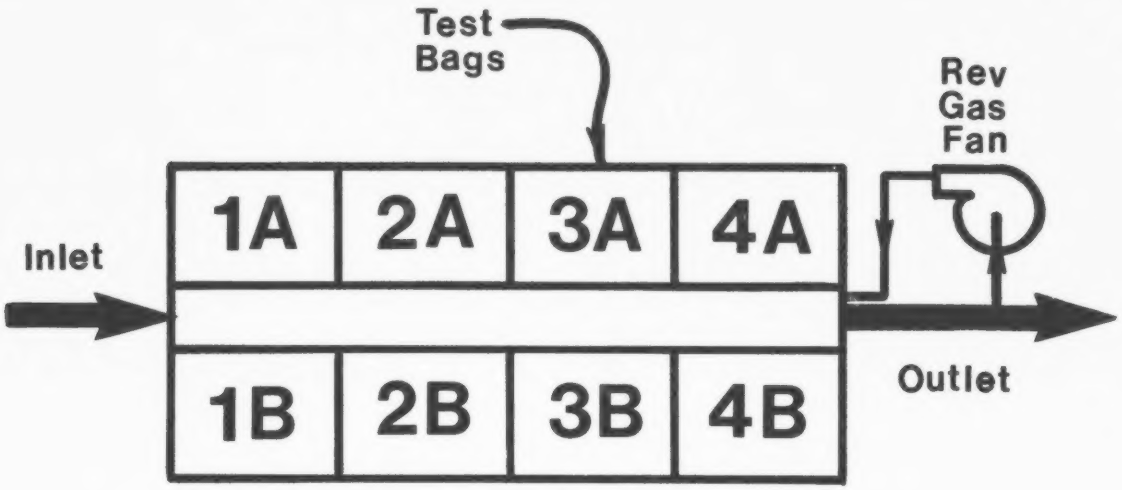


Figure 2. Plan-Fabric Filter

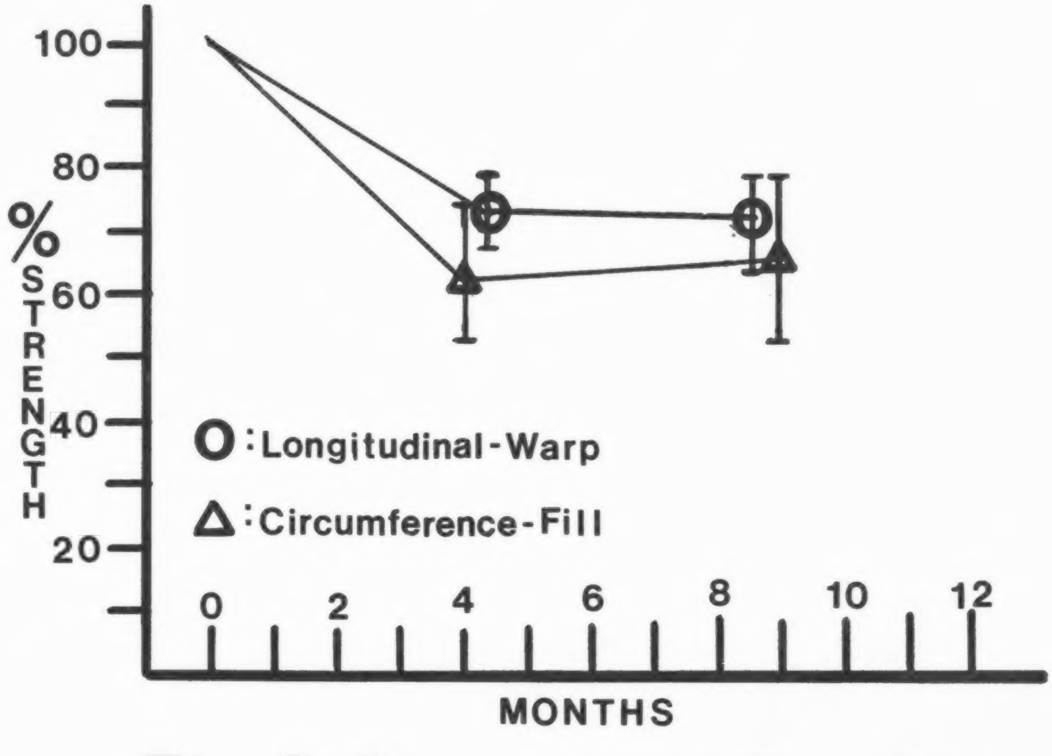
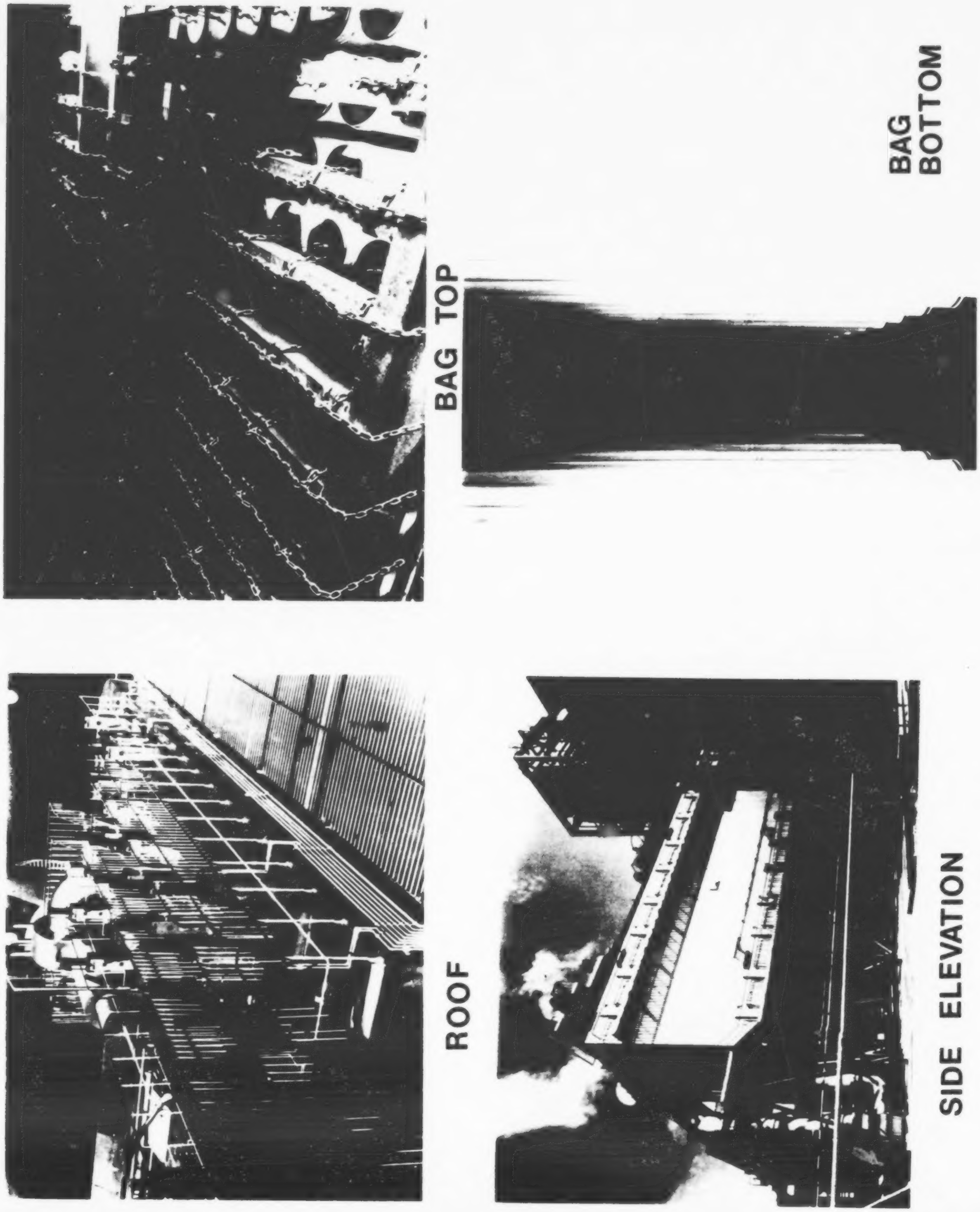


Fig. 3. Changes in Tensile



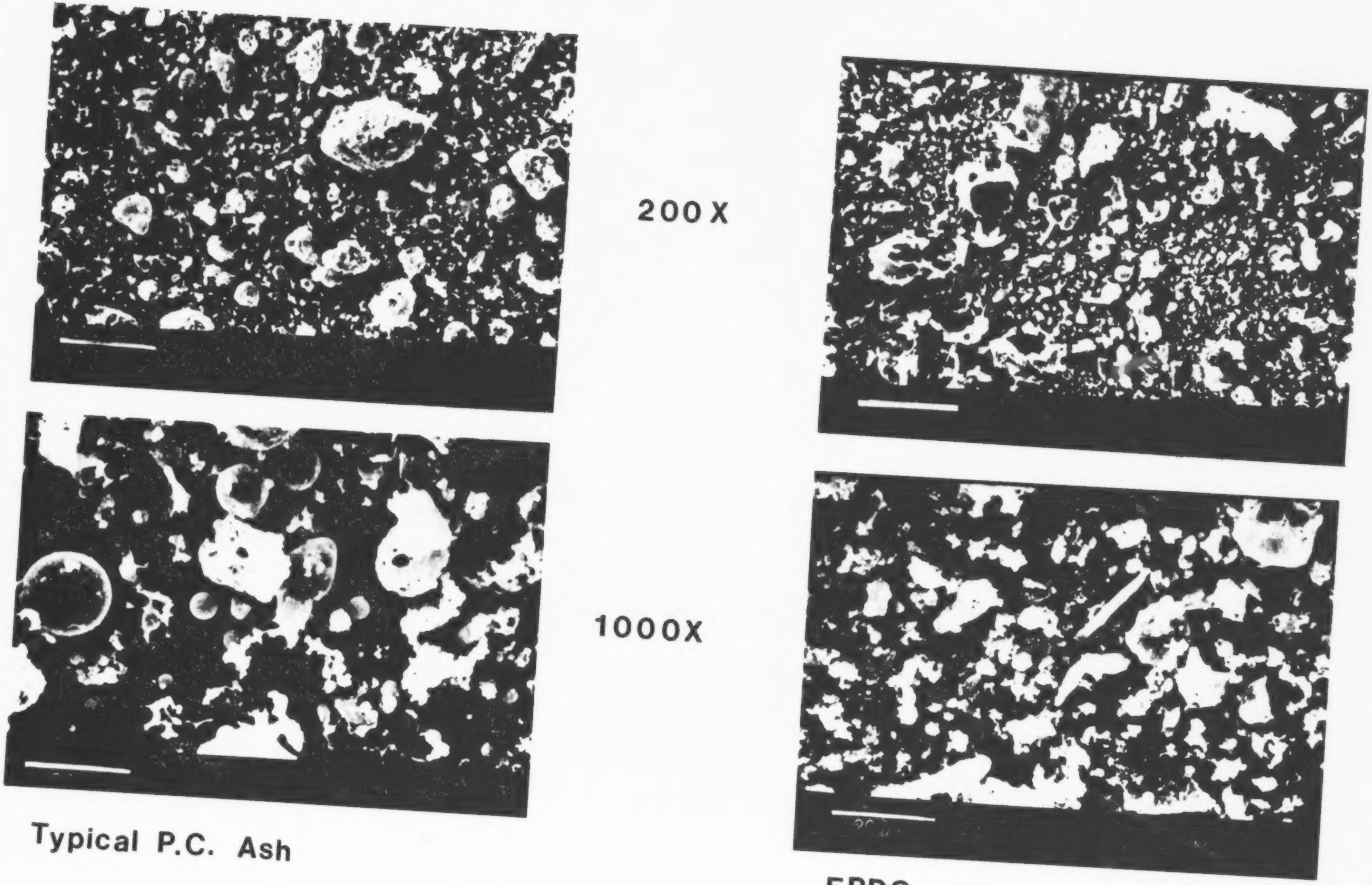
BAG TOP

BAG BOTTOM

ROOF

SIDE ELEVATION

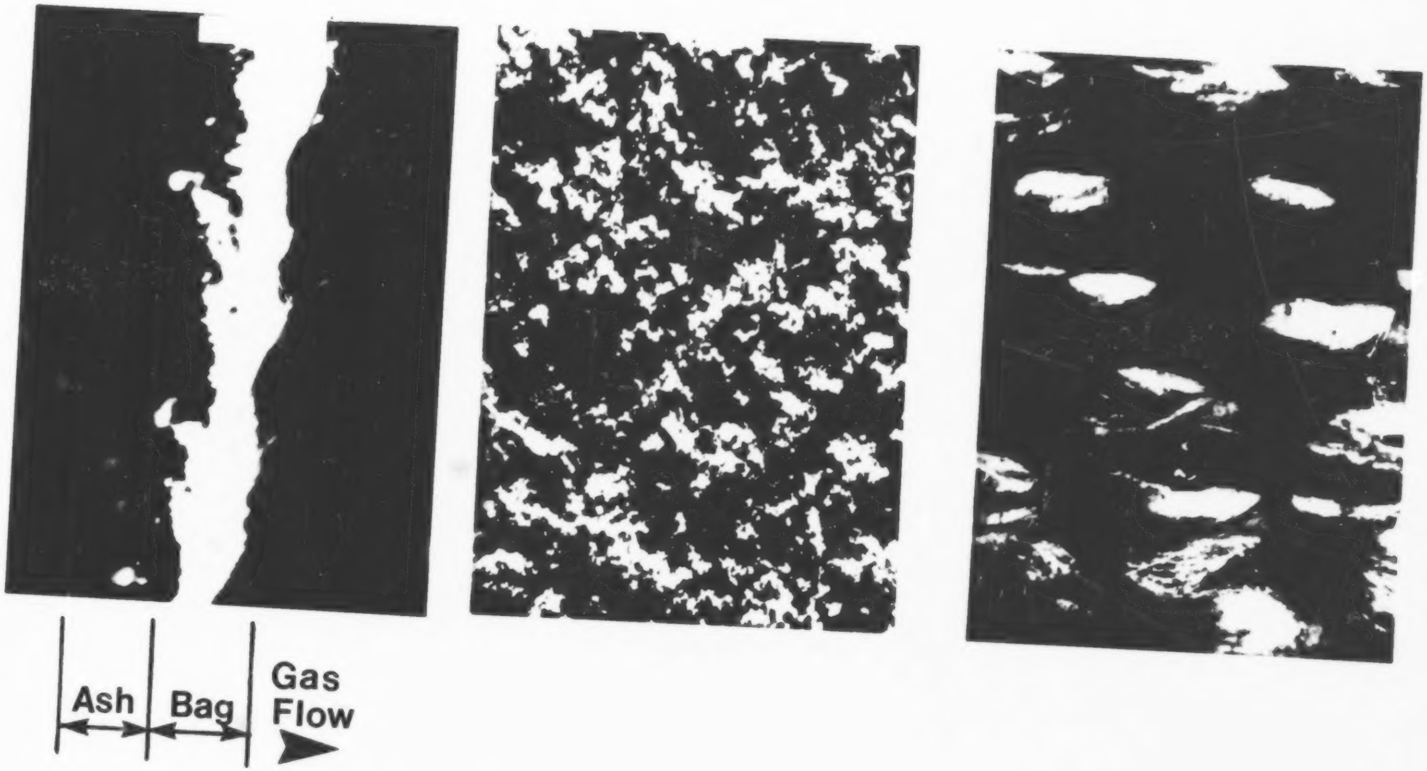
Fig. 4. Photographs



Typical P.C. Ash

EPDC AFBC Ash

Fig.5 Fly Ash

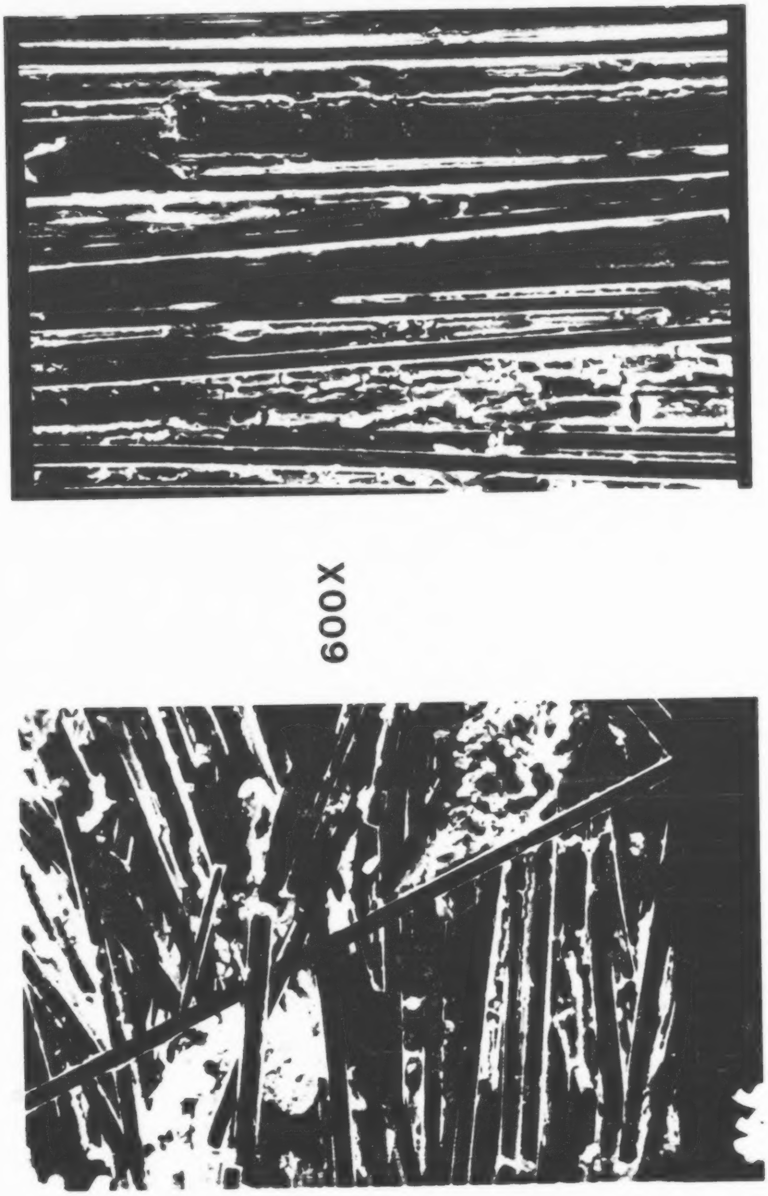


Cross Section

Ash Side

Clean Side

Fig.6. Fiberglass Bag 25X



600X

WARP

FILL

Fig. 7. Bag Yarns

Ash Surface 50X

Ash Surface 1000X

Ash Surface

Ash Surface

200X

View A

TEFLON-B

View B

MEMBRANE

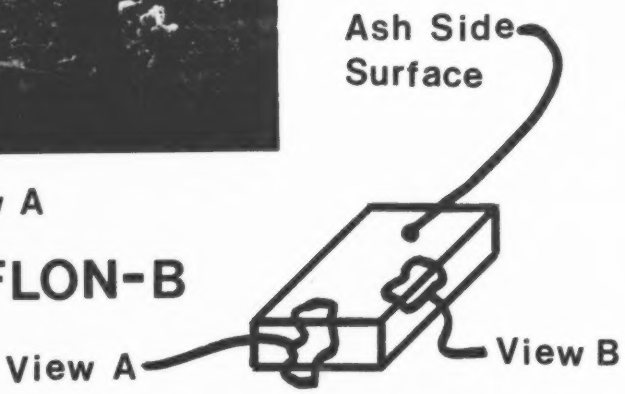
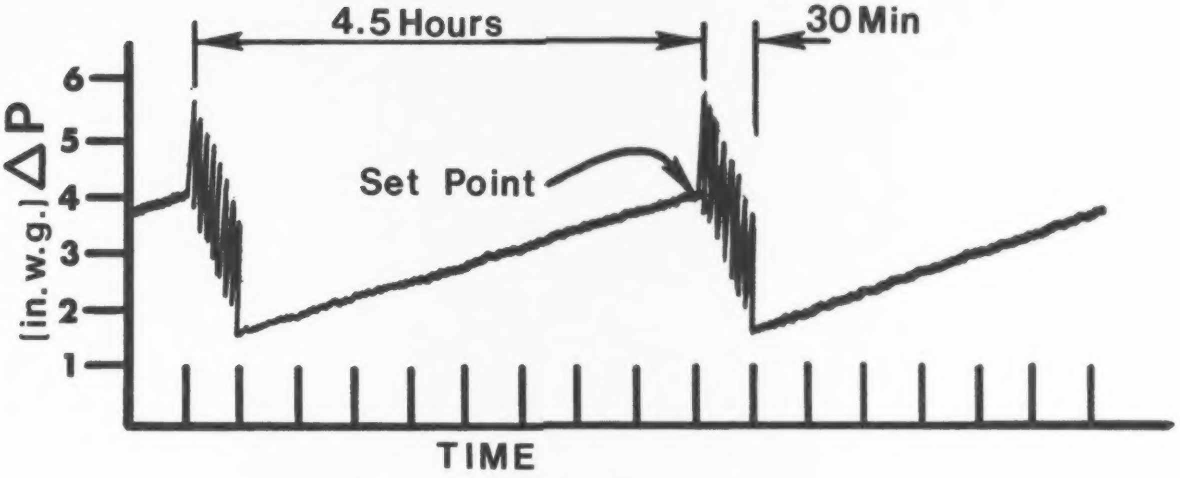


Fig. 8. Teflon-B/Membrane Bag



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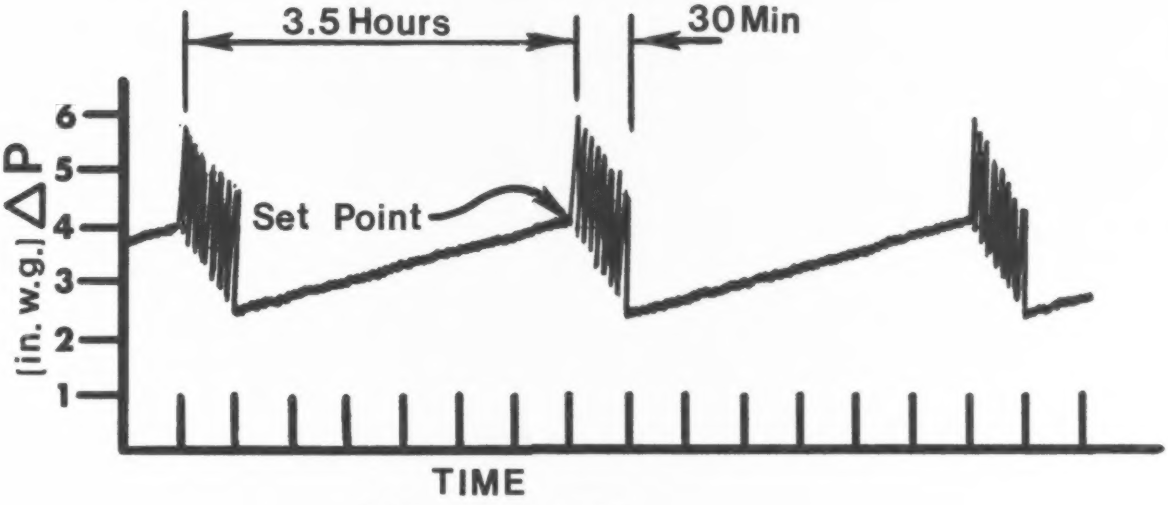

Nov. '87

Figure 9. Press. Recordings

TABLE 1 SPECIFICATION OF FABRIC FILTER

Flue Gas Flow Rate	162,000 ACFM
Flue Gas Inlet Temperature	266°F
Inlet Ash Concentration	8.11 gr/SCFD
Emissions at Fabric Filter Outlet	0.004 gr/SCFD
Pressure Loss, Flange to Flange	6 inches W.G.
Filter Bag Size	8 inch diameter
Filter Bag Type	Woven Glass (Teflon B)
Quantity of Compartments	8
Filter Speed Gross	1.44 FPM
Filter Speed Net-1	1.87 FPM
Filter Speed Net-2	2.18 FPM

TABLE 2 FILTER BAG SPECIFICATION

	Original	Test Bag A	Test Bag B
Filters	Woven Glass	Woven Glass	Woven Glass
Weave	3 x 1 Twill	3 x 1 Twill	3 x 1 Twill
Weight (oz./sq. yard)	10.3	10	14.8
Finish	10% Teflon B	Gore-Tex 3234	Micro-Tex 814F
Avg. Permeability (FPM) @ 1/2 inch W.G.	40	10	5

TABEL 3 COAL ANALYSIS

Coal	#1	#2	#3
Date Fired	10-86 to 3-87	4-87 to 5-87	6-87 to 12-87
Approx. Heating Value (BTU/LB)	11,800	11,700	12,500
Moisture % Weight	4.2	8.0	4.1
Ash % Weight	13.3	9.5	8.6
Sulfur % Weight	0.9	0.4	0.7
Volatile Matter %	31.6	25.8	28.2
Fixed Carbon %	50.0	56.3	59.1

TABEL 4 ASH ANALYSIS (% WEIGHT)

Coal	#1	#2	#3
CaO	12.2	17.8	10.9
$CaSO_4$	7.6	5.9	13.4
$CaCO_3$	5.1	5.6	5.3
SiO_2	35.8	35.4	28.0
Al_2O_3	22.6	16.8	10.9
Fe_2O_3	3.1	1.4	----
MgO	1.4	0.4	0.7
C	7.9	14.6	18.2
Other	4.3	2.1	12.6

TABLE 5 FABRIC FILTER TEST PROGRAM

Phase 1 (04/87 to 03/89) Investigations

A - Changes in Fabric Filter operating performance

- o Outlet emission rate
- o Rate of pressure loss increase and recovery of pressure loss by reverse air cleaning

B - Changes in properties of the filter cloth with the lapse of time (weight, perm, strength, etc.)

C - Performance relative to collecting sub-micron particulates

D - Equipment durability through periodical inspection of bags, dampers, walls, hoppers, doors, etc.

E - Problems during operation

Phase 2 (04/89 to 03/92) Improvement

A - Optimization of the cleaning cycle

B - Potential increase of filter speed

C - Upgrade to better performing type of filter bags

D - Application of a optimized control system

E - Understanding of the influence of different fuels

F - Understanding of the influence of combustor, SOx removal, and flue gas denitrification on the fabric filter

TABLE 6 FABRIC FILTER PERFORMANCE TESTING

Test Date	Feb. 1987	Nov. 1987
Flue Gas Flow Rate (ACFM)	155,000	152,000
Flue Temperature (°F) Inlet	316	280
Flue Temperature (°F) Outlet	304	270
Ash Concentration Inlet (gr/SCFD)	3.0	2.7
Ash Concentration Outlet (gr/SCFD)	Less than .004	Less than .004
Pressure Loss Before Cleaning (in. w.g.)	4.0	4.0
Pressure Loss After Cleaning (in. w.g.)	1.6	2.4
Particle Mass Mean Diameter, Inlet (u)	2.5	16
Avg. Time Between Fabric Cleaning (hr)	4.5	3.5
Coal Burned	No. 1	No. 3

TABLE 7 FILTER BAG TESTING (WOVEN GLASS)

Test Date	May 1987	Sept. 1987
Permeability (C.F./S.F./Min. @ 1/2" W.G.)	3.0	2.8
Elongation - Warp (%)	3.2	3.1
Fill (%)	3.3	3.7
Ash Deposits (oz./yd^2)	9	16
Tensile Strength - Warp (% Remaining)	72 Avg.	70 Avg.
Fill (% Remaining)	64 Avg.	65 Avg.

ANALYSIS OF THE PERFORMANCE OF AN ESP OPERATING DOWNSTREAM OF A SPRAY DRYER ON HIGH-SULFUR-COAL FLUE GAS

by

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ABSTRACT

This paper presents an analysis of data from the 10-MW Spray Dryer/Electrostatic Precipitator (ESP) Pilot Plant operated by the Tennessee Valley Authority at the Shawnee Station outside of Paducah, Kentucky. The pilot plant treats 35,000 acfm of flue gas from a wall-fired B&W boiler burning high-sulfur coal. The ESP is configured with four sections of collector plates, 23 ft. high and 9.2 ft. long, which provide a nominal specific collection area (SCA) of 400 ft^2/kacfm. A parametric study was undertaken to characterize the performance of an ESP operating downstream of a spray dryer flue gas desulfurization system. A comparison was made of the ESP collection efficiency, particle characteristics, and electrical operating parameters measured at baseline (flyash only) and spray dryer conditions. Tests were run at different SCA values by de-energizing electrical sections of the ESP. Tests were also run at different velocities to characterize the "non-ideal effects" of an ESP. The biggest impact of the spray dryer was an increase in inlet mass concentration by nearly a factor of 10. Although the reduced flue gas temperature produced improved ESP electrical operating conditions, these conditions did not result in a corresponding increase in the efficiency of the ESP. Analysis of the data indicates that re-entrainment may have limited the performance of the ESP.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

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INTRODUCTION

Due to plant life extension programs and potential acid rain legislation, utilities firing high-sulfur coal may be faced with retrofitting flue gas desulfurization (FGD) systems in existing plants. One potential cost-effective option is the spray dryer technology. Although spray dryers have been demonstrated at several full scale utility boilers, all of the existing systems are operating on low-sulfur flue gas. In addition, the majority of these installations use fabric filters downstream of the spray dryer whereas the high-sulfur retrofit candidates will most likely have ESPs. Therefore, there is insufficient data to evaluate the potential for high-sulfur spray dryer retrofit applications.

The TVA 10-MW Spray Dryer (SD)/ESP Pilot Plant Program is a three year project designed to obtain data on the retrofit application of spray dryer technology for medium-to-high-sulfur coal burning boilers. The program is co-funded by the TVA, EPRI, and Ontario Hydro. The test program is designed to address questions which must be resolved prior to commercialization of the high-sulfur SD/ESP applications. This program will provide a parametric investigation of the spray dryer, document the impacts on the ESP, and generate data on process design, operational problems, and hardware requirements.

This paper presents an analysis of data on the characterization of the ESP obtained during Phase III of the program. Phase III was the high-sulfur-coal test phase which was conducted between March and December, 1987. The test program was divided into three parts. A series of tests designated III-A were conducted to provide baseline data for the flyash-only condition. A parametric investigation was conducted by changing the gas velocity and the SCA of the ESP to cover a range of interests. A second series of tests, III-B, was conducted to characterize the SO_2 removal performance of the spray dryer/ESP system as a function of spray dryer operating variables. The parametric investigation involved an analysis of the effect of the following parameters: atomizer tip speed, inlet gas temperature, approach-to-saturation temperature, reagent ratio, and recycle rate. The third series, III-C, involved a repeat of the III-A tests with the spray dryer operating. These tests were conducted to evaluate the impact of the spray dryer on the ESP. The primary parameters were the ESP gas velocity and the SCA.

This paper analyzes the data from the III-A and III-C tests. The parametric investigation of the spray dryer will be reported at the First Combined FGD and Dry SO_2 Control Symposium to be held in St. Louis, MO on October 25-28, 1988.

BACKGROUND

For low-sulfur coal applications, spray dryers have proven to be an attractive alternative for flue gas desulfurization. There are currently fifteen U.S. utilities where spray dryers have been installed on full scale units with three more in the construction or planning stages. The performance of these units has been well documented (1-6). All fifteen of the operational systems are operating on coals with sulfur levels below 1.5% and the systems are designed for 60 to 90% SO_2 removal. Of these fifteen spray dryer systems, twelve have fabric filters, three have ESPs.

HIGH-SULFUR-COAL APPLICATIONS

While spray dryers have been successfully demonstrated for low-sulfur coal, this information is not directly applicable to the high-sulfur applications. Relative to the performance of an ESP downstream of a spray dryer, the primary difference between the two applications is that the high-sulfur installations will generally be designed for higher levels of SO_2 removal and will, therefore, have lower approach-to-saturation temperatures. The high-sulfur applications are also characterized by higher inlet temperatures. Both of these differences will require a greater amount of water to cool the gas. Since the performance of the spray dryer is optimized by maintaining a high percent solids slurry fed to the atomizer, the increased cooling water results in a proportional increase in solids loading to the flue gas. Another difference between the two applications is that the low-sulfur coals have a much higher alkalinity in the ash. Since the effects of these differences are not known, it is necessary to develop data on high-sulfur coal spray dryer/ESP systems.

There has been some previous research on spray dryer technology for high-sulfur-coal boilers performed at the pilot plant level, but it has primarily involved applications with fabric filters. EPRI has performed pilot plant studies at the Arapahoe Test Facility on a spray dryer with SO_2 spiked flue gas to simulate a high-sulfur coal application (7). EPRI and EPA have conducted high-sulfur-coal tests at the 100-MW demonstration system at the Northern States Power Riverside Station (8). Research is currently on-going at the EPRI High Sulfur Test Center using a 4-MW pilot spray dryer operating on the slipstream of a 643-MW boiler burning a 3% sulfur coal (9). One of the few programs involving an SD/ESP on high-sulfur coal was conducted by TVA using a 1.5-MW pilot plant at the Shawnee Test Facility (10).

SPRAY DRYER/ESP SYSTEMS

The performance of a SD/ESP system is of primary interest for the retrofit situation because the majority of the pre-NSPS boilers have ESPs. Since the operation of an ESP is very sensitive to particle loading and flue gas conditions, it is necessary to characterize the impact of a spray dryer on ESP performance. The majority of the operating data on SD/ESP systems has been obtained from the three full-scale SD/ESP systems. However, these systems are all on low-sulfur-coal boilers and the ESPs all have very large specific collection areas, 128 to 143 $m^2/m^3/s$ (650 to 725 $ft^2/kacfm$). The high-sulfur application will involve retrofits to older ESPs with a range of SCA from 35 to 90 $m^2/m^3/s$ (180 to 450 $ft^2/kacfm$).

The primary effect of the spray dryer on the ESP is the increased mass loading due to the sorbent material dispersed into the flue gas by the spray dryer and solid particles produced from conversion of SO_2. In order to maintain the baseline outlet emission levels, the collection efficiency of the ESP must be increased to capture the spray dryer material.

This increase in efficiency could be provided by improved ESP operating conditions resulting from cooling of the flue gas by the spray dryer. The cooling of the gas stream reduces the volume of flow treated by the ESP. Since the performance of an ESP is proportional to the specific collection area, the decrease in gas volume results in an increased SCA which should increase collection efficiency. The reduced flow rate also decreases the velocity in the ESP which could reduce some of the non-ideal effects such as re-entrainment and sneakage. In addition, the gas density is increased at the lower temperature which should increase the operating voltage of the ESP and hence improve the ESP performance. The decrease in viscosity with temperature will also reduce the drag on the particles and increase collection efficiency.

However, there are potential detrimentals effects on the performance of ESP due to the spray dryer. The increased loading might cause a space charge buildup in the ESP inlet field that can quench corona current and limit charging of particles. The magnitude of the space charge effect is proportional to the particle concentration, and if it is severe, it can reduce collection efficiency. In addition, the spray dryer material might lead to an increase in particle resistivity which is a key parameter affecting the performance of the ESP. Most high-sulfur ESPs rely on inherent conditioning of the ash by flue gas SO_3. Since the SO_3 is very reactive it will quickly be removed in the spray dryer which would destroy this conditioning effect. However, at the lower approach temperatures, the increase surface moisture on the particles should be sufficient to reduce the particle resistivity.

DESCRIPTION OF THE FACILITY

PILOT PLANT

The Spray Dryer/ESP Pilot Plant, located at the TVA Shawnee Steam Plant near Paducah, Kentucky, extracted a slip stream from the Number 9 boiler. This 150-MW wall-fired B&W boiler had a separate coal contract to fit the needs of the pilot plant. For the Phase III program a western Kentucky bituminous coal with a 4.0% sulfur content was burned. This provided SO_2 levels on the order of 2,500 ppm.

Figure 1 shows a schematic of the test facility. The flue gas was withdrawn downstream of the boiler air preheater and multiclone system at a nominal rate of 16.5 m^3/s (35,000 ft^3/min). The slipstream passed through a preheater/cooler heat exchanger which provided a mechanism to control the gas temperature to the pilot plant. A venturi, located upstream of the preheater, monitored the flow rate of the flue gas through the system. The flow rate was controlled by dampers located upstream of the spray dryer and upstream of the pilot plant ID fan. The operating conditions and the range of test conditions are presented in Table I.

The spray dryer vessel incorporated the side-exit design and used a single rotary disc made of 316L stainless steel. The disc operated with a direct drive and variable speed control. The slurry delivery system consisted of a fresh lime system and a recycle system. Fresh pebble lime was pneumatically conveyed from the delivery truck to a storage silo. During Phase III a Mississippi Lime Company high reactivity quick lime was used. The pebble lime was fed to the detention slaker as needed to prepare lime slurry which was pumped to storage. The lime slurry feed tank was refilled by pumping fresh lime slurry from storage.

Waste material collected in the spray dryer and ESP was removed by screw conveyors and discharged into a bucket elevator which emptied into a storage silo. The stored waste material was emptied to a recycle mix tank. Makeup water was added to the recycle mix tank to prepare batches of slurry which were pumped to the recycle feed tank and recycled to the spray dryer. Excess recycle slurry was pumped to waste. Slurry from the recycle feed tank and the fresh lime slurry were mixed in the feed line to the atomizer.

The ESP design was specified to provide data that would be appropriate for scale up. Therefore, the ESP was designed with a plate height of 7 m (23 ft), which is on the same scale as many existing units. The ESP was built with four electrical sections and eight gas flow passages. An empty section was located at the downstream end of the ESP to be retrofitted and put in service if needed. The corona electrodes were made of a stainless steel spiral wire mounted in a rigid frame configuration. The collecting electrodes were made of roll-formed plates spaced 25.4 cm (10 in.) apart. Both the collecting and the discharge electrodes were rapped by falling hammers on a rotating shaft. The ESP design data are summarized in Table 2. A microprocessor based system provided the high voltage control and the rapping sequence.

INSTRUMENTATION

A computerized data processing system was used to collect and process data from a variety of automated sensors and analyzers installed at key locations within the pilot plant facility. The SO_2 level was monitored by Dupont UV Photometric SO_2 analyzers located at the venturi inlet, SD inlet, ESP inlet, and ESP outlet. Temperatures, pressures, and oxygen levels were also measured at these points in order to correct the concentrations and flow rates for the effects of in-leakage and heat loss. RTDs were installed to measure the temperature and transducers provide the pressure measurements. Thermox zirconium oxide oxygen analyzers were used for the O_2 measurements. All of the slurry feed systems were automatically monitored and controlled by magnetic flow meters. The primary and secondary voltages and currents of each section of the ESP were also continuously monitored. The opacity was measured at the outlet of the ESP using a Lear Siegler transmissometer.

In addition, several manual methods were used to document the performance of the system. The mass concentration at the inlet and outlet of the ESP were measured using modified Method 17 sampling trains using in-stack filters operating at the flue gas temperature in the duct. These sampling trains also provide data on the moisture content of the flue gas.

Particle size distributions were also measured at the ESP inlet and outlet using University of Washington Mark III and Mark V in-stack impactors with Flow Sensors precutters. In order to provide the most meaningful particle size distribution data for analyzing the performance of the ESP, the impactor data was reduced to determine the physical diameter of the particles. This was accomplished by using the actual particle density which was measured with a Micromeritics Model 1503 Helium Pycnometer. Particle resistivity was measured using SoRI type in-situ resistivity probe. Voltage/current curves were measured manually every four hours. The mass loadings, particle size distributions, and particle resistivities were entered into the data processing system following the verification of the measurements.

TEST PLAN

In order to provide a data base on spray dryer/ESP systems that would be useful for a wide variety of retrofit applications, it is important to characterize the ESP as a function of SCA. In a parametric study the SCA can be varied by either changing the flow rate or by de-energizing sections. The problem with changing the flow rate is that it also changes the gas velocity which has other effects in an ESP unrelated to SCA. A large portion of the emissions from a high performance ESP are due to "non-ideal effects" such as re-entrainment, sneakage, and rapping. These are all very strong functions of the gas velocity. Because of this, when the flow rate is varied it is not possible to discriminate the change in performance due to the change in SCA from that due to the change in velocity. Therefore, it was decided that when SCA was the primary parameter being varied, the SCA would be changed by de-energizing electrical sections of the ESP. De-energizing sections also reduces the number of electrical sections in the ESP to 2 or 3, which is more typical of the smaller SCA ESPs.

Besides determining the effect of SCA, it is also important to characterize the effects of velocity on re-entrainment. This can be determined by comparing the results from two tests conducted at an identical SCA but at different velocities. Theoretically these two tests should provide the same performance because of the identical specific collection area. Therefore, any difference in the results should be directly attributable to the effect of velocity.

A test plan was developed to satisfy the above requirements which involved five different test conditions that are summarized in Table III. The plan incorporated three different values of SCA at the design flow rate of 16.5 m^3/s (35,000 ft^3/min) which provided a range of SCA from 38 to 76 $m^2/m^3/s$ (193 to 386 $ft^2/kacfm$). Two of the tests involved a change in velocity to characterize the non-ideal effects. Test 5 has the same SCA as test 1 but the velocity is reduced 25%. Although, the SCA values for 2 and 4 are not identical, they are close enough to evaluate the effect of an increased velocity. The SCA values could not be matched because the pressure drop through the pilot plant system limited the maximum flow rate to 20.8 m^3/s (44,000 ft^3/min).

For Tests 2, 3, and 5 the upstream section(s) were de-energized. Although this provides an open plenum for settling of larger particles, these particles would settle in the ESP anyway or would be easily precipitated and therefore do not significantly affect the overall performance. On the other hand, if the last sections were de-energized, many of the charged particles exiting the previous sections would still be collected because, despite the absence of an externally applied electric field, the electric field produced by the space charge would cause the charged particles to migrate outwardly toward the plates.

The performance of the ESP was first characterized for the baseline (i.e. flyash only) condition at an inlet temperature of 160^oC (320 ^oF). For each test, the ESP was operated at constant conditions for several days while sufficient measurements were made to document the performance. Each test included 1 to 2 days of lineout operation to achieve steady-state operation at the new condition. During the lineout period, VI curves and outlet mass loading measurements were made to determine when conditions stabilized.

The test block was then repeated to determine the performance of the ESP under spray dryer conditions. In order to reflect operation in a retrofit mode, the gas flow rates were set at the inlet to the spray dryer. This allows the ESP to take advantage of the reduced flow due to the cooling of the gas. In this manner it is possible to determine if the improved operating characteristics due to the lower flow rate, lower velocity, higher density, lower resistivity, and lower viscosity are sufficient to overcome the increased mass loading.

During this test period the spray dryer was operated under constant conditions. The inlet temperature to the spray dryer was maintained at 160°C (320 °F). The reagent ratio was set at a value of 1.3 moles of calcium in the fresh lime slurry per mole of SO_2 in the flue gas. Maximum recycle was used to control the solids content of the slurry to the wheel. The slurry feed rate was set to provide an approach-to-adiabatic-saturation temperature of 10 °C (18 °F)

RESULTS

The values for mass loading, particle size distribution, and resistivity that are presented here represent the averages of a minimum of three separate measurements. Most of the particle size data are plotted as differential mass distributions because this provides a graphical means to compare distributions with different mass loadings. The area under each curve is proportional to the amount of mass in each particle size interval. These plots combine the relative size distribution information from the impactors and the mass concentration data from the filter tests.

BASELINE CONDITIONS

Following startup and shakedown of the pilot plant, the baseline tests were conducted to characterize the ESP at flyash only conditions. The average flue gas characteristics for the baseline tests are summarized in Table IV. Except for the inlet mass loadings, the conditions are typical of high-sulfur coal flue gas with the particle resistivity in a range which provides near optimum ESP operation (10^{10} ohm-cm). The inlet mass loadings were lower than most boilers because of the multiclones located upstream of the pilot plant. The multiclones not only reduce the mass loading, but they bias the size distribution by primarily removing larger particles.

The first series of measurements was the warranty test at the ESP design conditions which was designated as Test 2-A-01. Figure 2 shows the secondary voltages and current characteristics for these tests. The particle resistivity provided for excellent operating conditions with average electric field strengths greater than 3 kV/cm and current densities of 64 nA/cm^2, which are typical of a well operating ESP, in all but the first section. The V/I curve for the first section is offset due to the space charge produced by the higher particle loading. The performance of the ESP under these conditions is summarized in Table V. As would be expected, the ESP collected the flyash at a very high efficiency, 99.8%, which met the design performance warranty.

Particle size measurements were made at the inlet and outlet of the ESP in order to determine the collection efficiency as a function of particle size. Figure 3 is a plot of the penetration as a function of the particle physical diameter. The increase in penetration as the particle size decreases is expected because the theoretical efficiency of an ESP is proportional to the particle diameter.

Two sets of tests were then run to determine the effect of a change in SCA at the design flow rate. In Test 3-A-01, one section was de-energized to reduce the SCA by 25%. Although the efficiency dropped to 99.7%, the ESP was still providing excellent performance. In Test 3-A-04 a second field was de-energized which reduced the SCA by 50% from the design condition. This resulted in a substantial decrease in efficiency to 98.5%. However, the outlet particle concentration still met the State of Kentucky regulation of 0.11 lb/MMBtu.

The results of the particle size distribution measurements made at the outlet of the ESP are plotted in Figure 4 for the three values of SCA. The curves show a general trend of increasing emissions in the 2 to 10 micrometer range with decreasing SCA. Since the majority of mass at the inlet to the ESP was located in this particle size range, the emissions were most likely due to insufficient collector plate area rather than re-entrainment.

If re-entrainment was a problem, it should increase at higher velocities. Two tests were conducted to determine the effects of velocity on the performance. Test 3-A-02 was set up with the same SCA as Test 3-A-01 but the velocity was increased from 1.16 to 1.46 m/s (3.8 to 4.8 ft/s). For both of these tests the efficiency was 99.7% indicating that there were no detrimental non-ideal effects due to the increased velocity. Test 3-A-03 provided a similar comparison with Test 2-A-01 at a reduced velocity. However, rather than the performance increasing or remaining the same for the lower velocity, the efficiency decreased to 99.6%. This reduction in efficiency was probably due to the decreased inlet loading for Test 3-A-03 because the outlet emissions are nearly identical.

SPRAY DRYER CONDITIONS

Following a seven month parametric study of spray dryer operation, the ESP was characterized for spray dryer conditions. The average flue gas characteristics for this series of tests are summarized in Table VI. As expected, the primary differences in the gas stream characteristics are that the mass increased by nearly a factor of 10 and the gas was cooled from 158.3°C (317 °F) to 62°C (144 °F). Also the moisture increased from 7% to 13% and the cooling of the gas provided sufficient conditioning of the flyash to reduce the resistivity to 7×10^9 ohm-cm.

The increased gas density produced by the gas cooling was expected to improve electrical operating conditions, but the increased loading could produce corona quenching. The effects of the spray dryer on electrical operation are demonstrated by comparing the VI characteristics of Test 2-A-01 with Test 3-C-01 which are plotted together in Figure 5. The characteristics for only the first and fourth fields are shown. A comparison of the first fields shows the effects of both the increased mass loading and the lower temperature. The changes in the final section would be solely due to the gas temperature effect as the majority of the particles were collected upstream.

The curves in Figure 5 show improved electrical operation was achieved in both the upstream section and the downstream section. The electric field strength increased by almost 30% in electrical section 1 and 20% in section 4. It should be noted that if the operating points were not limited by the 200 mA power supply, the improvement might have been even greater. The increased mass loading did not produce a space charge problem because most of the mass is in the larger size intervals.

During the spray dryer tests, approximately 80 to 90% of the mass was collected in the precutter. In order to characterize these larger particles, a Bacho analysis was performed on a sample of material collected in a thimble filter at the inlet to the ESP. Figure 6 is a plot of the cumulative distribution of the inlet particle size showing the data from four impactor runs and the Bacho analysis. There was excellent agreement between all of the data.

Figure 7 shows a comparison of the inlet size distributions for baseline and spray dryer conditions. Most of the increase due to the spray dryer material was in the particle size range greater than 4 micrometers. The decrease in mass below 2 micrometers might be due to the difficulty in obtaining sufficient mass on the lower stages of the impactor before overloading the upper stages. However, it could indicate some agglomeration of the flyash in the spray dryer.

Table VI is a summary of the performance of the ESP for spray dryer conditions. These conditions were designed to duplicate the baseline tests except for the reduced gas flows due to the cooling of the gas which resulted in lower velocities and higher values of SCA. The data shows that the collection efficiencies were very similar to those measured in the baseline tests.

This is somewhat surprising because the efficiency should improve at spray dryer conditions. The increased loading was primarily in the larger size ranges, and therefore, should have been easier to collect. In addition, at spray dryer conditions the ESP operated at field strengths 20 to 30% higher than the base case and the viscosity was reduced 20%. The combination of these factors should have provided a significant increase in overall ESP collection efficiency.

The outlet particle size distributions were analyzed to determine why the collection efficiencies were not higher. Figure 8 is a comparison of the outlet distributions of Tests 3-C-01 and 2-A-01. In the range of 2 to 5 micrometers, the mass emissions were on the order of 2 to 20 times higher in the spray dryer case. This is quite unusual as Figure 7 shows that the spray dryer produced very little change in the loading in this size range.

Figure 9 shows a comparison of the ESP collection efficiency as a function of particle size for the spray dryer and baseline conditions for the particle size range of 1 to 15 micrometers. This range was selected because, as shown in Figure 8, it represents the majority of the emissions. It would be expected that since the spray dryer conditions produced higher operating voltages and a lower viscosity, it should have had a higher efficiency. For particles greater than 8 micrometers, this was true as Test 3-C-01 had a greater efficiency (i.e. less penetration) than Test 2-A-01. However, in the range of 1 to 5 micrometers the penetration was 5 to 10 times greater for the spray dryer case. This indicates that the problem may be due to a non-ideal effect, such as sneakage or re-entrainment. Since, the configuration of the ESP was identical in both cases and the velocity was slightly lower in the spray dryer case, it is unlikely that sneakage would be increased at spray dryer conditions. Therefore, the problem was most likely some characteristic of the sorbent/flyash mixture that made it more easily re-entrained.

Another indication that re-entrainment was occurring can be seen in the comparison of the results from Tests 3-C-02 and 3-C-03 in Table VI. If re-entrainment were indeed the source of the problem, it would be enhanced at higher velocities. When the velocity was increased from 1.08 to 1.2 m/s (3.54 to 3.94 ft/s) the emissions nearly doubled even though the SCA was 15% higher in the higher velocity case. This indicates that re-entrainment was indeed a problem in the ESP at spray dryer conditions.

Reentrainment could be due to either particles dislodged during rapping or the scouring of the particles from the collector plates by the gas flow. Tests conducted by EPRI on full scale boilers(11) have shown that rapping puffs have a mass median diameter of 6 micrometers, which is similar to the size distribution measured in this program. However, because the opacity meter was operating on a 0-100% range, there is no real-time data available to differentiate between the rapping and non-rapping reentrainment.

Figure 10 presents the outlet size distributions for the three tests at the same flow rate but with different numbers of sections energized. The distribution for Test 3-C-02 with three fields is very similar in shape to Test 3-C-01 with most the emissions due to particles in the 2 to 10 micrometer range. However, the big change in going to 2 fields, Test 3-C-05, occurs in particles in the size range of 5 to 20 micrometers. This curve indicates that the two section ESP had insufficient SCA to collect the material added by the spray dryer. However, the performance of the three section configuration was high enough that re-entrainment was a significant factor.

DISCUSSION AND CONCLUSIONS

The particle and flue gas characteristics produced by the spray dryer provided for excellent ESP operating conditions. The decreased temperature provided a significant increase in ESP electric field strength over baseline conditions. Because of the large size of the spray dryer material, even a factor of 10 increase in mass loading did not produce a space charge problem in the inlet field. In addition, the cooling and moisture increases were sufficient to condition the particulate matter to lower the resistivity values to the optimum ESP operating range. Resistivity measurements made during the spray dryer parametric study showed that even at an approach temperature of 21 ^oC (38 ^oF), the resistivity level was down to 4×10^{10} ohm-cm.

However, in spite of the improved operating conditions and the increase in the amount of easily precipitated material, the ESP efficiency did not increase at the spray dryer conditions. It appears that there was a re-entrainment problem at the spray dryer conditions. This is demonstrated by the increase in percent penetration of 2 to 5 micrometer particles by as much as a factor of 10 even though the electric field strength increased by 20 to 30% under the spray dryer conditions. In addition, there was a significant increase in emissions due to a velocity increase.

This problem must be resolved because the current data indicates that it would take an ESP with an SCA of 50 $m^2/m^3/s$ (400 ft^2/kacfm) just to meet a 0.1 lb/MMBtu standard for the spray dryer conditions. This would severely restrict the number of potential applications of the retrofit spray dryer technology. However, if the reentrainment can be reduced by adjusting the rapping or providing a conditioning agent to improve particle cohesion, the ESP performance would be greatly improved.

The problem with re-entrainment will be investigated during future testing at the TVA spray dryer/ESP pilot plant. Understanding this problem will provide a major step toward a solution that would result in a potentially cost effective alternative for retrofit flue gas desulfurization.

ACKNOWLEDGEMENTS

The authors would like to acknowledge the efforts of several individuals responsible for the management of the pilot plant, the day to day operation of the facility, and the processing of the data. This includes the following TVA personnel: H.B. Flora II, Dick Runyan, Russell Robards, Steve Newton, Ed Puschaver, and the TVA Shawnee 10-MW operations and maintenance support staff. We would also like to acknowledge Mr. Charlie Brown, the Radian on-site engineer, for his assistance with the data. Mr. Timothy Ebner of ADA Technologies produced all of the plots used in this paper.

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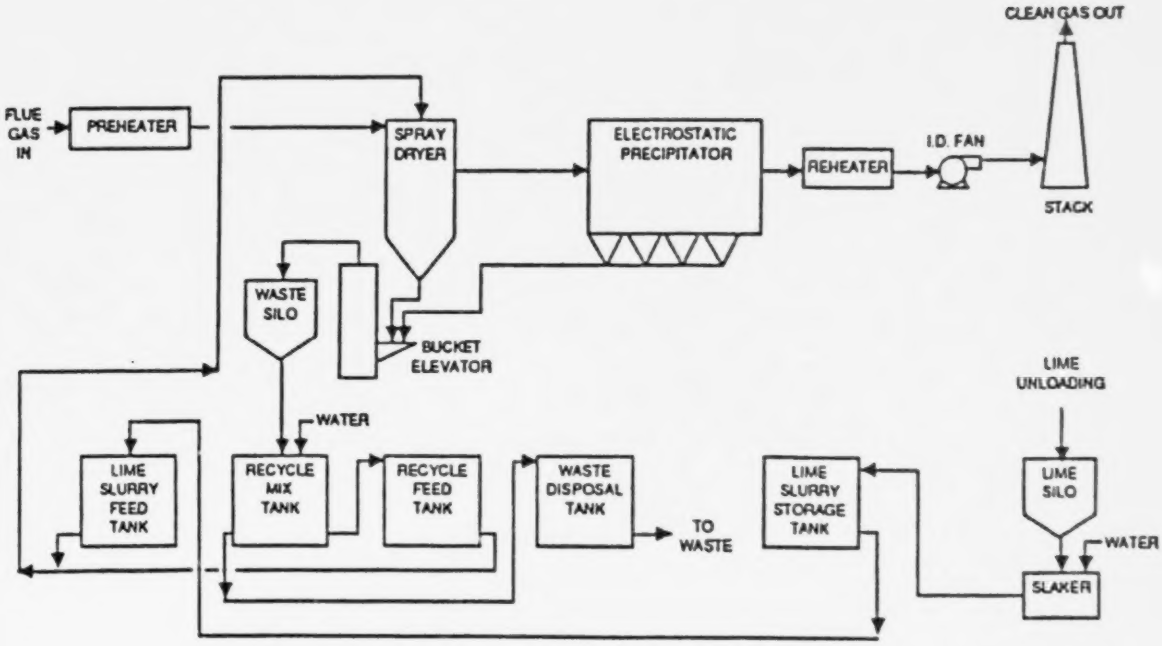

Figure 1. Process Flow Diagram for the TVA 10-MW Spray Dryer/ESP Pilot Plant.

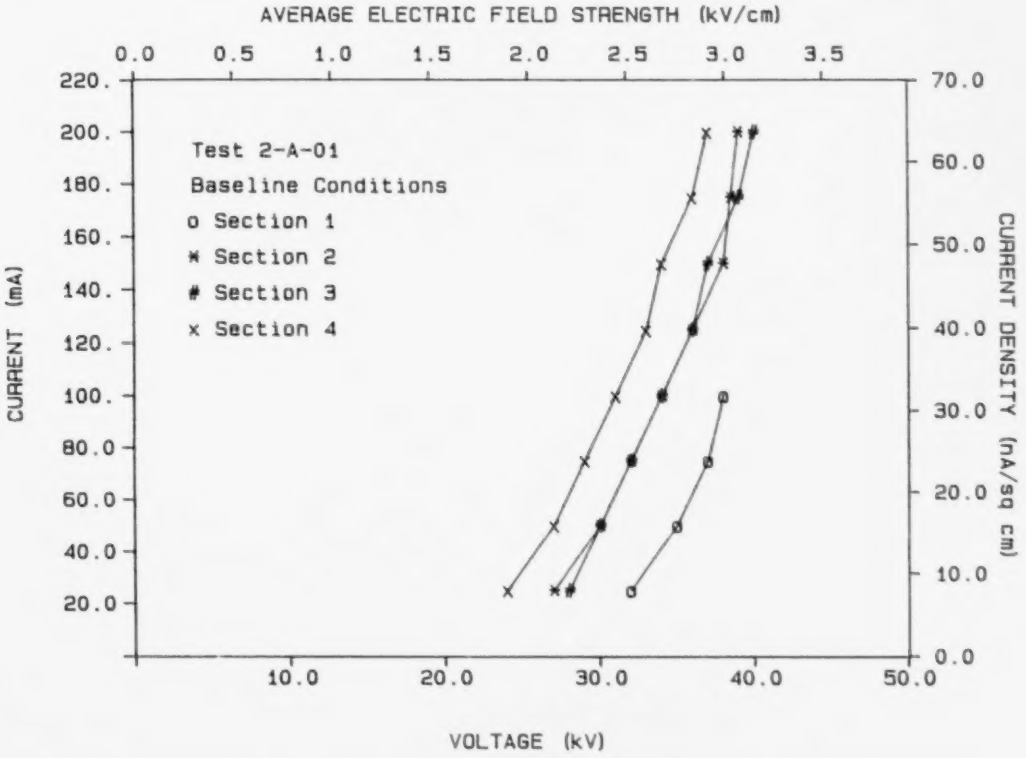

Figure 2. ESP Voltage and Current Characteristics for the Baseline Conditions.

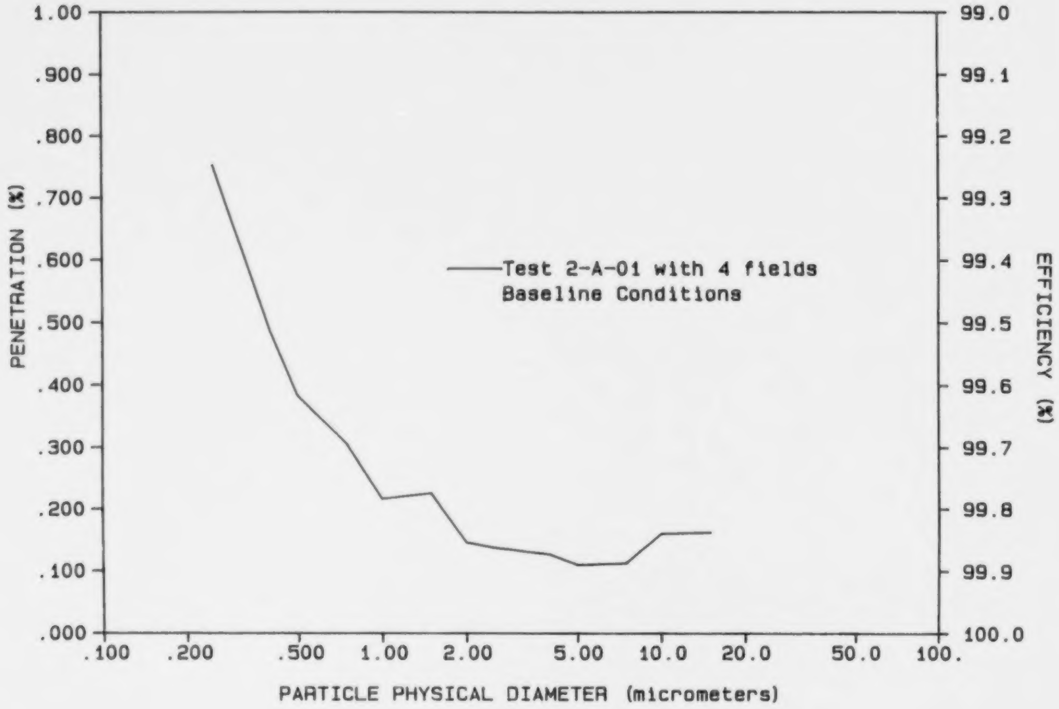

Figure 3. ESP Penetration and Collection Efficiency as a Function of Particle Diameter.

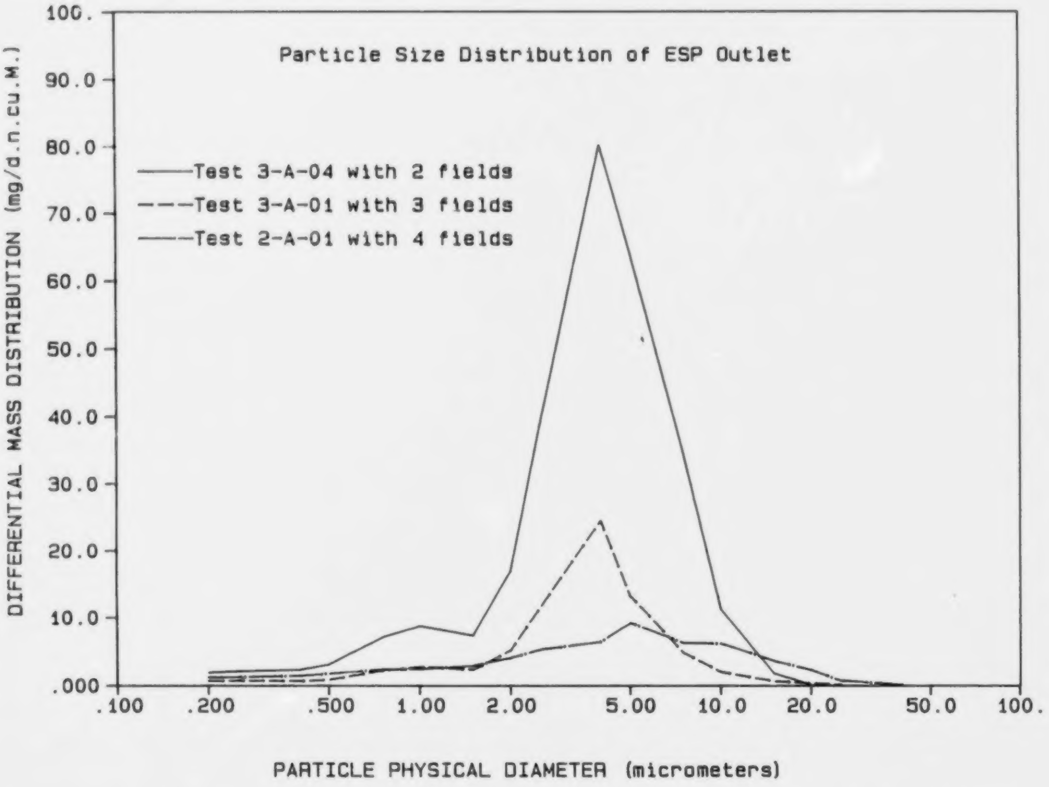

Figure 4. Outlet Differential Particle Size Distribution vs. SCA at Baseline Conditions.

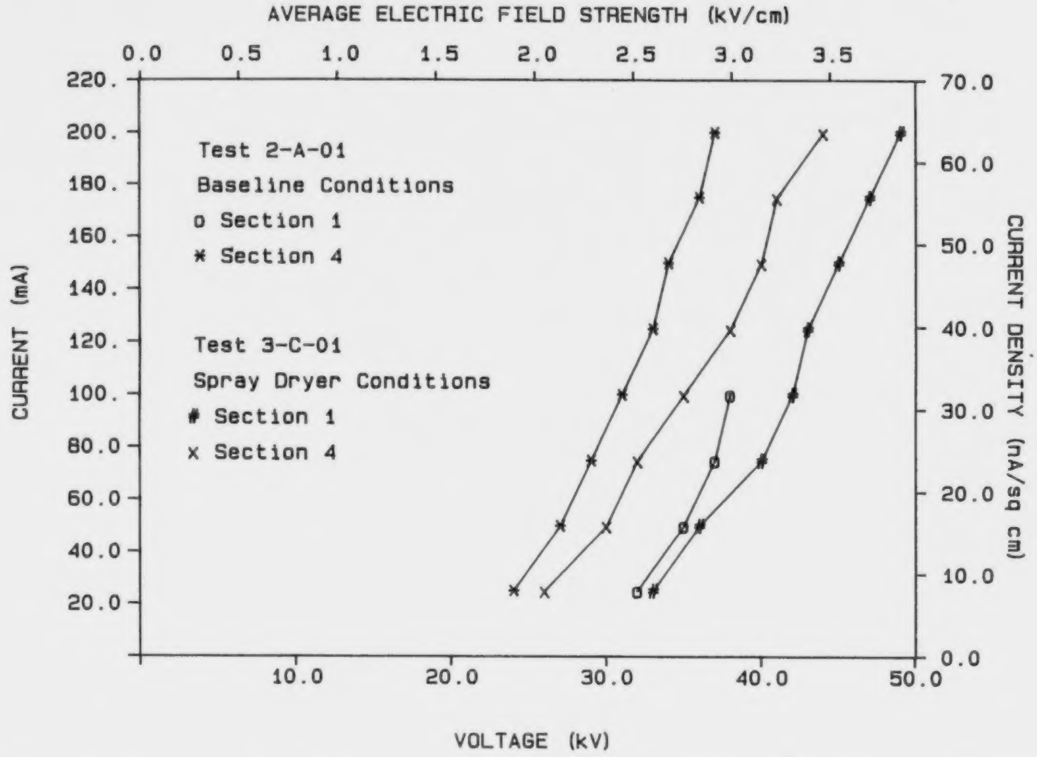

Figure 5. Comparison of Voltage and Current Characteristics at Baseline and SD Conditions.

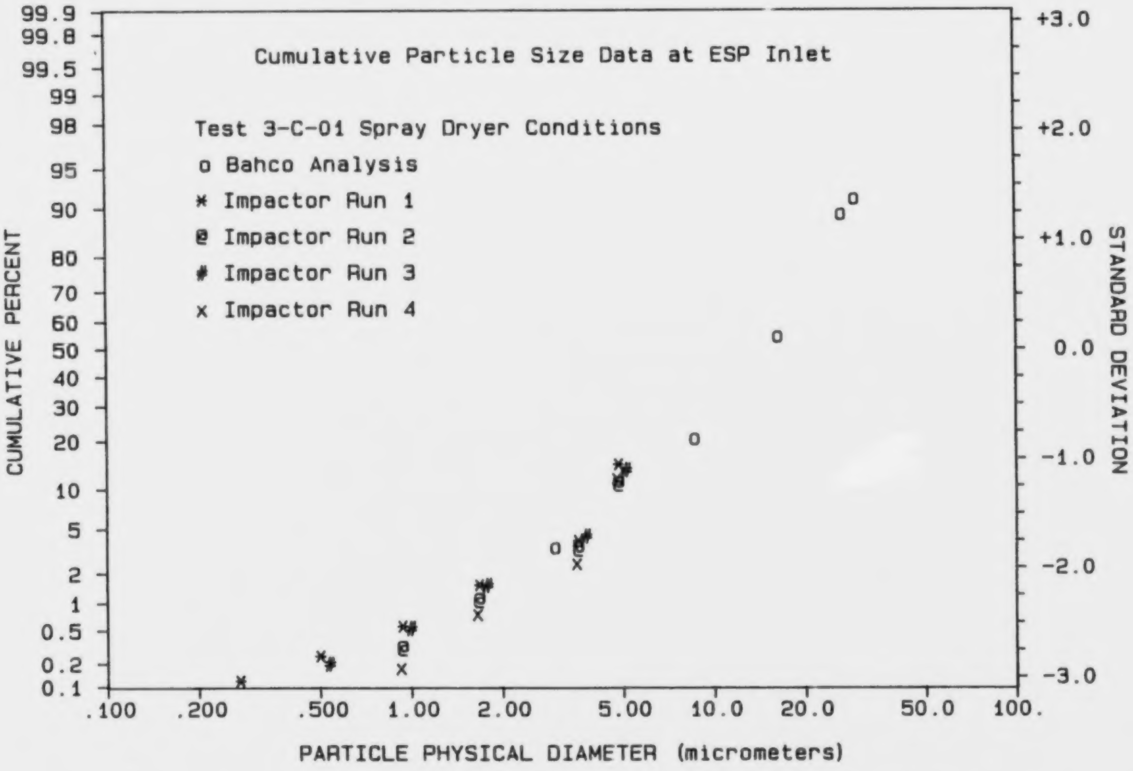

Figure 6. Cumulative Particle Size Distribution at the ESP Inlet for SD Conditions.

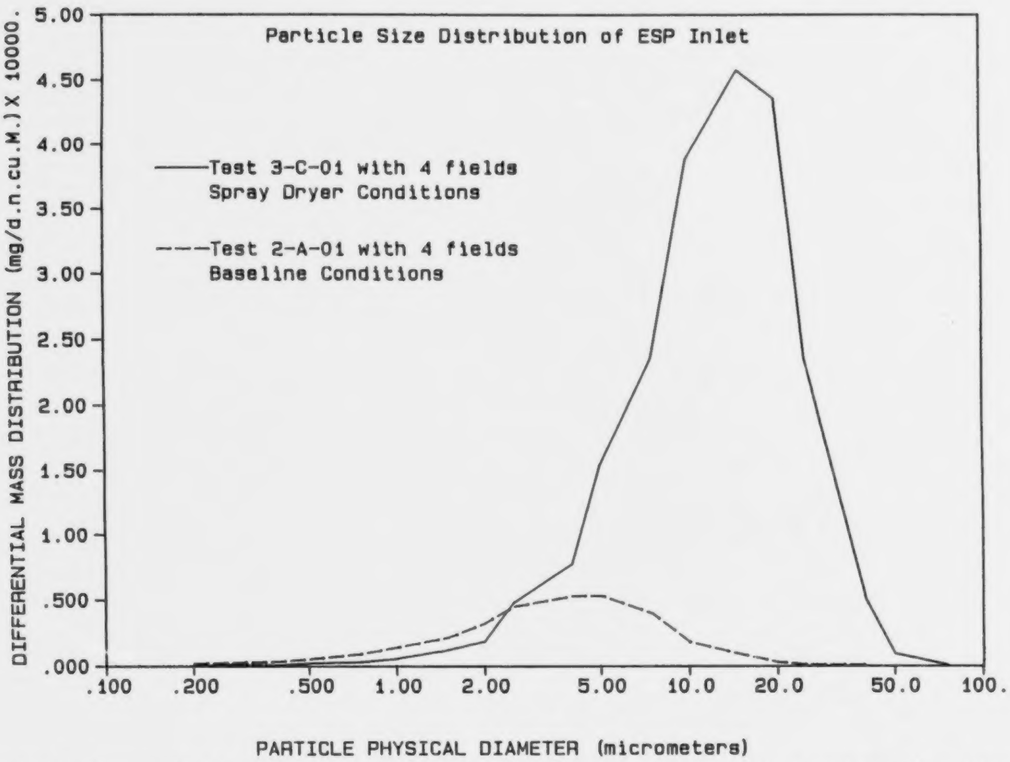

Figure 7. Comparison of Inlet Particle Size Distributions for Baseline and SD Conditions.

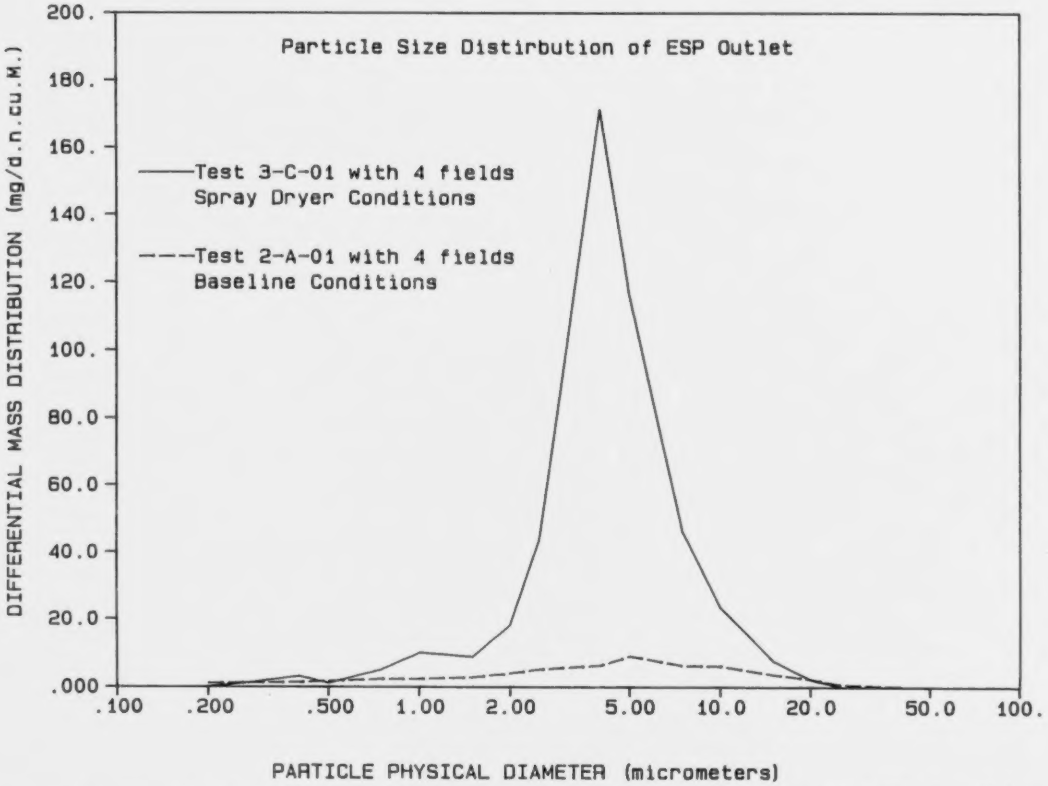

Figure 8. Comparison of Outlet Particle Size Distributions for Baseline and SD Conditions.

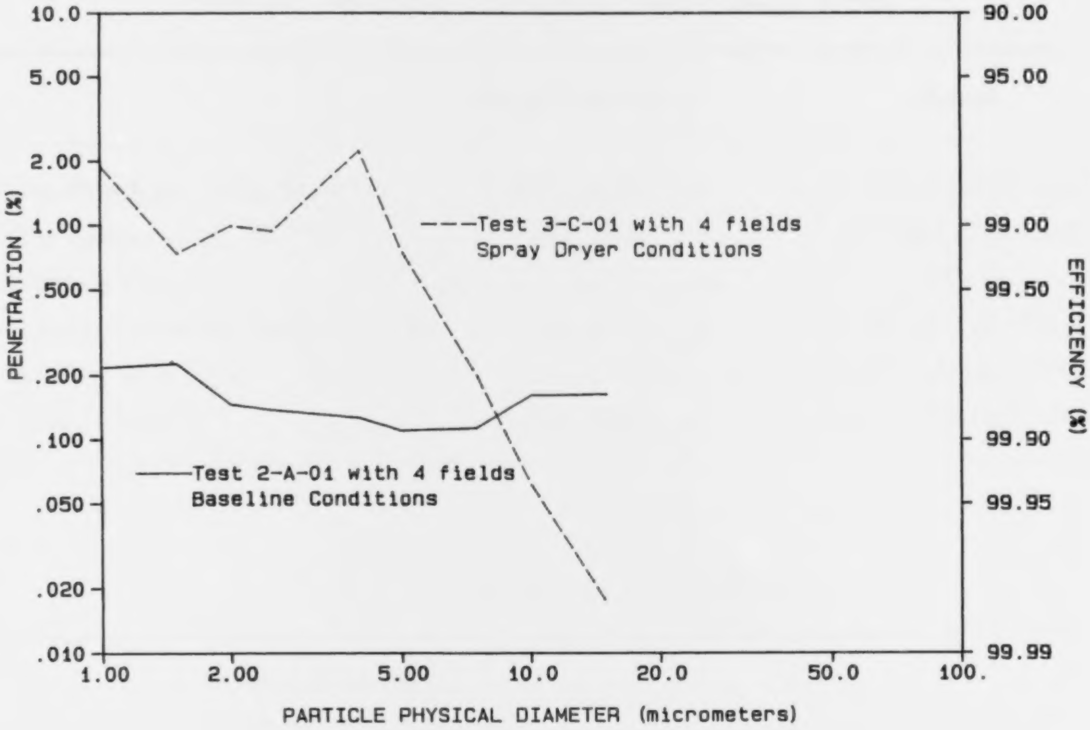

Figure 9. Comparison of Penetration vs. Particle Diameter for Baseline and SD Conditions.

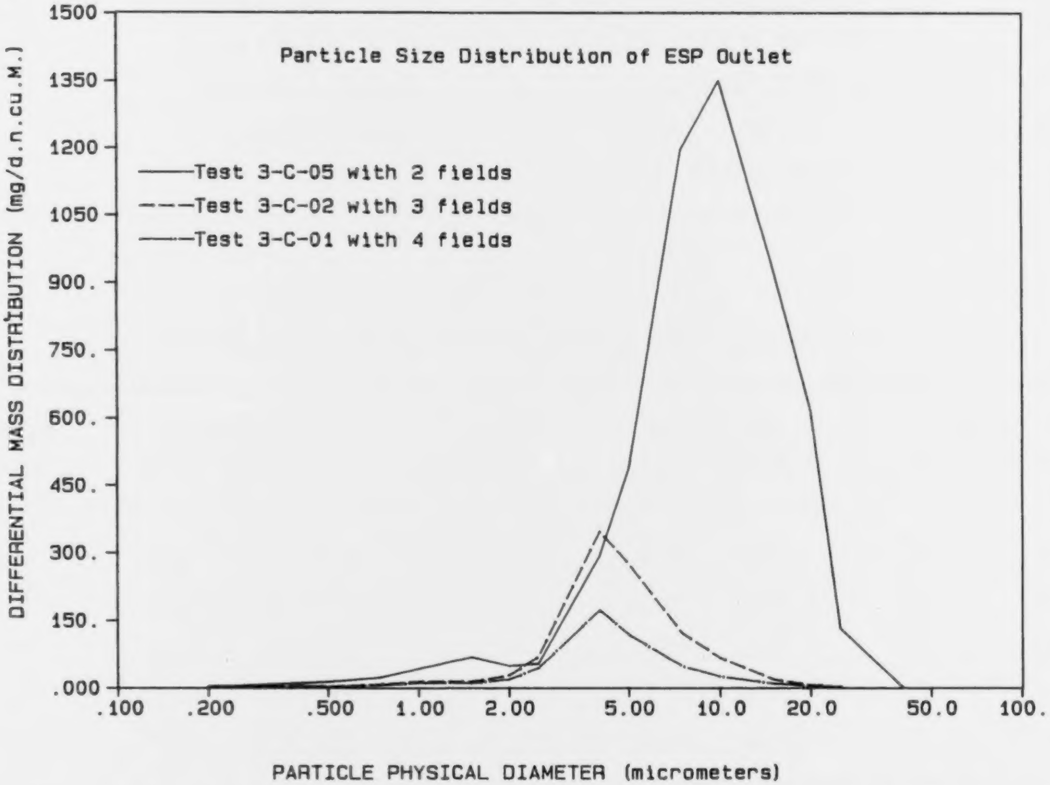

Figure 10. Outlet Differential Particle Size Distribution vs. SCA at SD Conditions.

TABLE I. PILOT PLANT OPERATING CONDITIONS.

	Design Conditions	Range
Inlet Gas Flow Rate, m^3/s (acfm)	16.5 (35,000)	11.8 - 20.8 (25,000 - 44,000)
Inlet Gas Temperature, °C (°F)	160 (320)	127 - 160 (260 - 320)
SD Residence Time, s	10	8.5 - 12
Approach-to-Saturation Temp., °C (°F)	11 (20)	10 - 21 (18 - 38)
ESP Velocity, m/s (ft/s)	1.15 (3.8)	0.82 - 1.5 (2.7 - 4.8)
SCA, s/m (ft^2/kacf)	76.0 (386)	30.2 - 107.0 (153 - 541)

TABLE II. ESP DESIGN VARIABLES

Number of Sections	4
Number of Gas Passages	8
Plate spacing	25.4 cm (10 in.)
Plate Height	7 m (23 ft)
Plate Length	2.8 m (9.2 ft)
Plate Area/Section	314 m^2 (3,382 ft^2)
Corona Electrode	Spiral
Power Supplies	55 kV, 200 mA

TABLE III. TEST PLAN FOR ESP CHARACTERIZATION

TEST NO.	FLOW RATE m^3/s	(acfm)	ESP SECTIONS	VELOCITY m/s	(ft/s)	SCA s/m	(ft^2/kacf)
1	16.5	(35,000)	4	1.15	(3.8)	76	(386)
2	16.5	(35,000)	3	1.15	(3.8)	57	(290)
3	16.5	(35,000)	2	1.15	(3.8)	38	(193)
4	20.8	(44,000)	4	1.5	(4.8)	60	(307)
5	12.3	(26,000)	3	0.8	(2.8)	76	(386)

TABLE IV. AVERAGE FLUE GAS CONDITIONS FOR THE ESP CHARACTERIZATION.

TEST BLOCK	INLET MASS g/dnm^3	(gr/dnft^3)	ESP TEMP. °C	(°F)	H_2O (%)	RESISTIVITY (ohm-cm)	SO_2 (ppm)
Baseline (Phase III-A)	3.14	(1.37)	158	(317)	7.0	2.8×10^{10}	2420
Spray Dryer (Phase III-C)	30.2	(13.2)	62	(144)	13.7	7×10^9	2270

TABLE V. SUMMARY OF ESP PERFORMANCE FOR BASELINE CONDITIONS

TEST NO.	FLOW RATE m^3/s	(acfm)	VEL. m/s	(ft/s)	FIELDS	SCA s/m	(ft^2/kacf)	INLET CONC. gr/acf	OUTLET CONC. gr/acf	EFFIC. %
2-A-01	16.3	(34,614)	1.1	(3.8)	4	77	(390)	1.151	0.00198	99.82
3-A-01	16.5	(34,961)	1.2	(3.8)	3	57	(290)	0.942	0.00240	99.73
3-A-02	20.7	(43,965)	1.2	(4.8)	4	60	(307)	0.727	0.00221	99.69
3-A-03	12.5	(26,564)	0.9	(2.9)	3	75	(381)	0.584	0.00220	99.60
3-A-04	16.6	(35,124)	1.2	(3.8)	2	38	(192)	0.646	0.00942	98.48

TABLE VI. SUMMARY OF ESP PERFORMANCE FOR SPRAY DRYER CONDITIONS

TEST NO.	FLOW RATE m^3/s	(acfm)	VEL. m/s	(ft/s)	FIELDS	SCA s/m	(ft^2/kacf)	INLET CONC. gr/acf	OUTLET CONC. gr/acf	EFFIC. %
3-C-01	15.5	(32,832)	1.1	(3.6)	4	81	(412)	9.223	0.0155	99.84
3-C-02	15.4	(32,623)	1.1	(3.6)	3	61	(311)	8.987	0.0304	99.66
3-C-03	17.8	(37,752)	1.2	(4.1)	4	70	(358)	10.286	0.0617	99.40
3-C-04	11.4	(24,242)	0.8	(2.6)	3	82	(419)	9.167	0.0256	99.73
3-C-05	15.0	(31,856)	1.1	(3.5)	2	42	(212)	8.58	0.219	97.46

E-SO_x SYSTEM FOR COMBINED PARTICULATE AND SO_2 CONTROL

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ABSTRACT

E-SO_x is a proposed retrofit system that makes use of an existing ESP for combined sulfur dioxide (SO_2) removal and particulate collection. Results of research conducted in the EPA 1000 acfm (28 m^3/min) ESP to establish baseline data and performance criteria for the E-SO_x system are presented. The results show the magnitude of the influence of lime, introduced as a slurry into the entrance section which is devoid of wires and plates, on the SO_2 concentration in the gas and on the collection of particulate in the remainder of the ESP. The reported work was done as part of a larger project with an objective of evaluating E-SO_x at large pilot scale at a plant site burning 2.5-3% sulfur coal. The majority of the small pilot experiments described were carried out at temperatures and SO_2 concentration to simulate plant site conditions. Within these limitations the paper describes the efforts to establish optimum process parameters for maximum SO_2 removal while maintaining acceptable ESP performance.

This paper has been reviewed in accordance with the U.S. Environmental Protection Agency's peer and administrative review policies and approved for presentation and publication.

INTRODUCTION

E-SO_x is a retrofit process, which combines electrostatic precipitator (ESP) technology for collecting particles and spray dryer technology for SO_2 removal in a single unit, for coal-fired boilers. The process makes use of a modified existing ESP with an auxiliary system for preparation and injection of a sorbent slurry into the ESP. E-SO_x is similar to some other proposed retrofit lime or limestone dependent processes, in that a dry solid waste is produced, which is collected along with the fly ash in the particulate collector, usually an ESP. The concept of E-SO_x appears to offer an attractive option for acid rain mitigation. An economic evaluation of E-SO_x, based on reasonable rates of SO_2 removal and lime utilization, has indicated that it can be cost effective.(1) The work that is reported here is from the first experiments to verify assumptions used in predicting the cost of removal and to examine the reproducibility of the process. Both the economic study and initial report on E-SO_x from this laboratory (2) gave impetus to a large pilot plant evaluation sponsored by the U.S. Environmental Protection Agency (EPA) and the Ohio Coal Development Office to begin in late 1988.

E-SO_x CONCEPT

In the E-SO_x retrofit concept shown schematically in Figure 1, the front end of an existing ESP is converted to a spray chamber where contact is made between gaseous SO_2 and wetted lime particles. Evaporation of the water also takes place in this converted section of the ESP so that the reaction product, excess lime, and fly ash which enter the remainder of the ESP are sufficiently dry for efficient ESP operation. The electrical components of the remainder of the ESP are left intact so that it operates as a particle collector. To provide this contacting chamber with a sufficient space to allow a finite drying time of 1 second or greater, the ESP will most likely lose 25-30% of its collector plate surface. At the same time the particulate load will increase by a factor of three or more. This particulate, a large fraction of which is calcium based, has an extremely high electrical resistivity. The sorbent reaction with SO_2 will eliminate sulfur trioxide (SO_3) effects that would tend to lower this resistivity. On the other hand, lowering of the gas temperature by the water spray will tend to more than compensate for resistivity increases due to ash composition. The E-SO_x concept, from the beginning, has recognized the availability of advanced ESP technology to aid in maintaining good performance in particulate collection. Cold pipe precharging and large diameter electrodes, for example, can be introduced in retrofitting ESPs to compensate for less collector surface or changes in resistivity characteristics.

The E-SO_x concept raises two fundamental questions which can only be answered by experiment. The first question regards the feasibility of removing substantial SO_2 from the rapidly moving gas by contact with slurry droplets and drying the droplets all within the space of one ESP section. The second question has to do with maintenance of an acceptable level of ESP performance under a reduced collector area and an increased particulate loading. This paper is a progress report on obtaining favorable answers to these questions.

EQUIPMENT

As shown in Figure 2, the experimental setup in EPA's Air and Energy Research Laboratory consists of the ESP with electrical components removed from the first section and a venturi scrubber to remove excess SO_2 before the gas is exhausted to the atmosphere. Figure 3 shows the layout of the 1000 acfm ESP. Outside air is heated to 149°C (300°F) prior to the injection of SO_2 to the desired concentration (usually 1500-2000 ppm) to simulate burning of moderately high sulfur coal. Fly ash is aspirated countercurrently into the simulated flue gas stream just before the cocurrent injection of the lime slurry. Slaked lime, containing 15-20% solids, is forced through a Casterjet two-fluid nozzle designed to provide an oblong spray pattern in the 1.27 m (50 in.) high x 0.38 m (15 in.) wide contacting chamber. The droplets, with a Sauter Mean diameter of 30 μm, have a 1.5 to 2 s residence time in the chamber to evaporate most of the water. Evaporation effects a temperature drop, the magnitude of which depends upon the volume of water sprayed. The spray chamber is made up of the entrance section and the first of the four ESP sections. The electrical configuration of the ESP is flexible, but for these tests, two cold-pipe prechargers were employed: one in the connecting space between the first and second sections, and one between the third and fourth sections. Conventional wire-plate electrodes were assembled in sections 2 and 4. In effect, this reduced the 25 s/m (128 SCA) ESP to a 12.5 s/m (64 SCA) ESP plus the prechargers. SO_2 monitoring is at the ESP exit just before the gas goes through the scrubber. A section of the chart recorder on the SO_2 monitor is reproduced in Figure 4 to show the pattern of removal. Following the chart movement from right to left, initially the gas measures approximately 2000 ppm SO_2. When the lime spray is activated, the SO_2 concentration drops rapidly to level off at 900 ppm. At the end of the test, water spray replaces the slurry and the SO_2 concentration returns to a level below the initial value. This demonstrates the wall effects of this unit. Unreacted lime, deposited on walls and fixtures, reacts with the SO_2. These effects will vary, but will reduce in magnitude as the size of ESP is increased in scale-up.

EXPERIMENTAL

The primary objectives were to determine the critical E-SO_x parameters and to find the best set of conditions for both SO_2 removal and particulate collection. This paper is a report of the progress made in meeting these objectives. Other tests, designed specifically to aid in scaling up to the large pilot evaluation, will be reported in future papers.

A number of tests were performed in which the concentration and rate of injection of the slaked lime spray were varied. In essence, looking at these two variables permitted an examination of the effect of two critical parameters: the approach to saturation temperature of the gas stream, and the calcium to sulfur ratio in the slurry/gas mix. In this series of experiments other parameters were fixed or held constant including the volumetric air flow rate, SO_2 concentration, initial temperature of simulated flue gas, type of spray nozzles, and the ESP configuration. The fixed conditions are tabulated in Table 1.

The minimum approach temperature depends on the amount of water that can be tolerated in the ash by the first electrical section of the ESP. In effect then, the minimum approach temperature is determined by the maximum water content of the ash which will not interfere with collection and handling of the particulate in the ESP. Once this is established the problem is reduced to finding the optimum stoichiometric ratio and the slurry concentration that will provide the correct combination of both calcium and water. The work is summarized in Figure 5. The correlation presented is between SO_2 removal and the approach temperature at several stoichiometric ratios. At extremely low approach temperatures, good removal is obtained at very low ratios. However, moisture in the ESP ash limits the lowest approach temperature to 15.5-16.7°C (28-30°F). Stoichiometric ratios of 1.3 to 1.4 produced a removal of 50% or better at a 16.7°C (30°F) approach to saturation. Results indicate a marginal improvement in SO_2 capture at ratios above 1.4, but the increase in ratio with a smaller return makes the higher ratios less cost effective.

The second objective was to determine the effect of the increased particle loading on the ESP. Results of a thorough quantitative analysis are not yet available, but it can be stated that, in the search for optimum SO_2 removal conditions, the ESP performance remained high. Efficiencies of 99% and above were obtained even in the face of a 4-fold increase in particulate loading and a 50% reduction in ESP SCA. The high efficiency is attributable to several factors taken together. Evaporation of water during the drying step makes E-SO_x a relatively low temperature process and, as a result the high electrical resistivity of lime sorbent/fly ash mixtures, is not encountered in E-SO_x. Figure 6 shows how the resistivity of E-SO_x particulate drops to low levels below 65°C (150°F). As a consequence, back corona is not a problem. There are other reasons for good ESP performance. A large fraction of the lime particulate and the reaction product particulate are relatively large particles. They tend to fall out in the spray chamber or to be collected in the first stage of the ESP, presenting no removal burden on the ESP. Another significant contribution to good performance may be in the use of a cold-pipe precharger (CPP). The CPP was not needed to perform its primary function of lowering the temperature of the dust layer collected on the cooled surface; however, in addition to its primary function, the CPP serves as an efficient collector of low resistivity dust in contrast to the small space it occupies in the ESP. For the E-SO_x process the latter may be the main advantage of using a CPP in the collector.

SUMMARY

Basic concepts of the E-SO_x process have been verified at small pilot scale with a simulated flue gas. Removal of 50% or more of the SO_2 emitted from a boiler burning a 2.5-3% sulfur coal appears to be a reasonable goal. At the level of lime usage required and the observed performance of the short-end ESP, experimental evidence appears to verify assumptions made in predicting a cost effective process.

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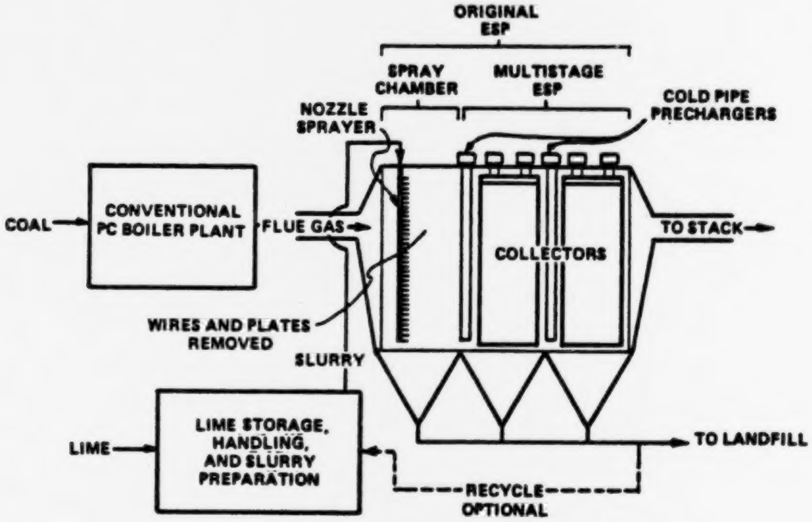


Figure 1. Schematic diagram of E-SO_x process.

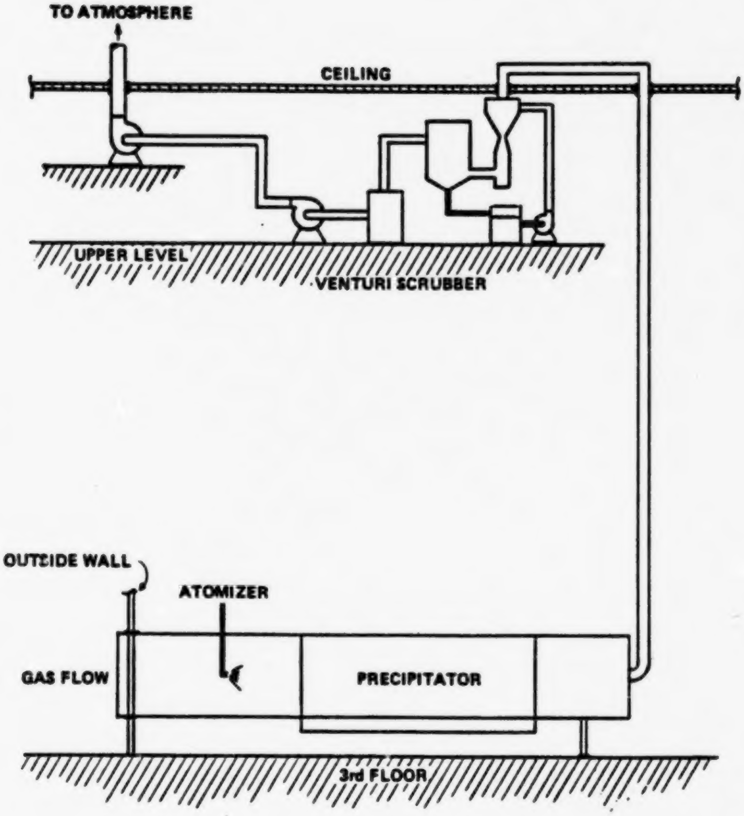
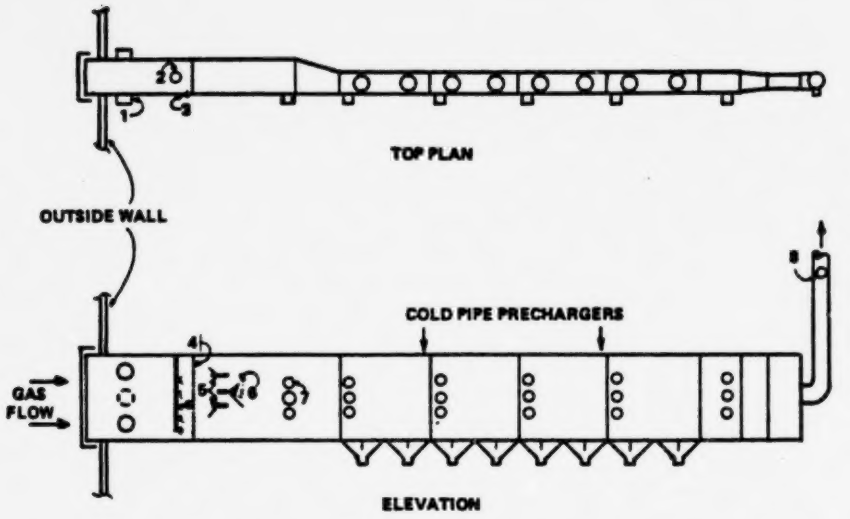


Figure 2. EPA E-SO_x pilot facility schematic .



1. NATURAL GAS BURNERS
2. SO_2 INJECTORS
3. BURNER TEMP. CONT. T.C.s
4. PERFORATED PLATE
5. ASH INJECTORS (AGAINST GAS FLOW)
6. ATOMIZER TIP
7. SAMPLING PORTS
8. OUTLET SAMPLE TAP

Figure 3. EPA E-SO_x pilot facility, 1000 acfm (28 m^3/min) ESP with spray zone.

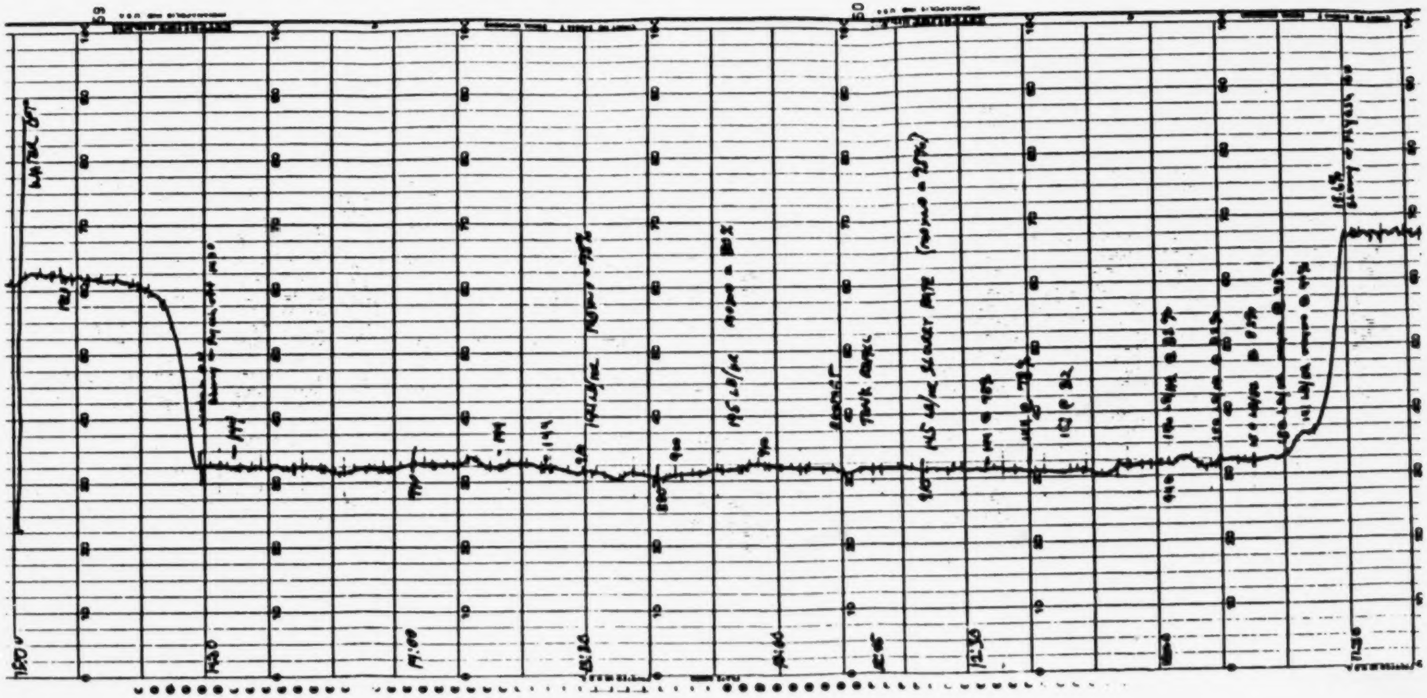


Figure 4. SO_2 removal by E-SO_x as recorded on SO_2 monitor.

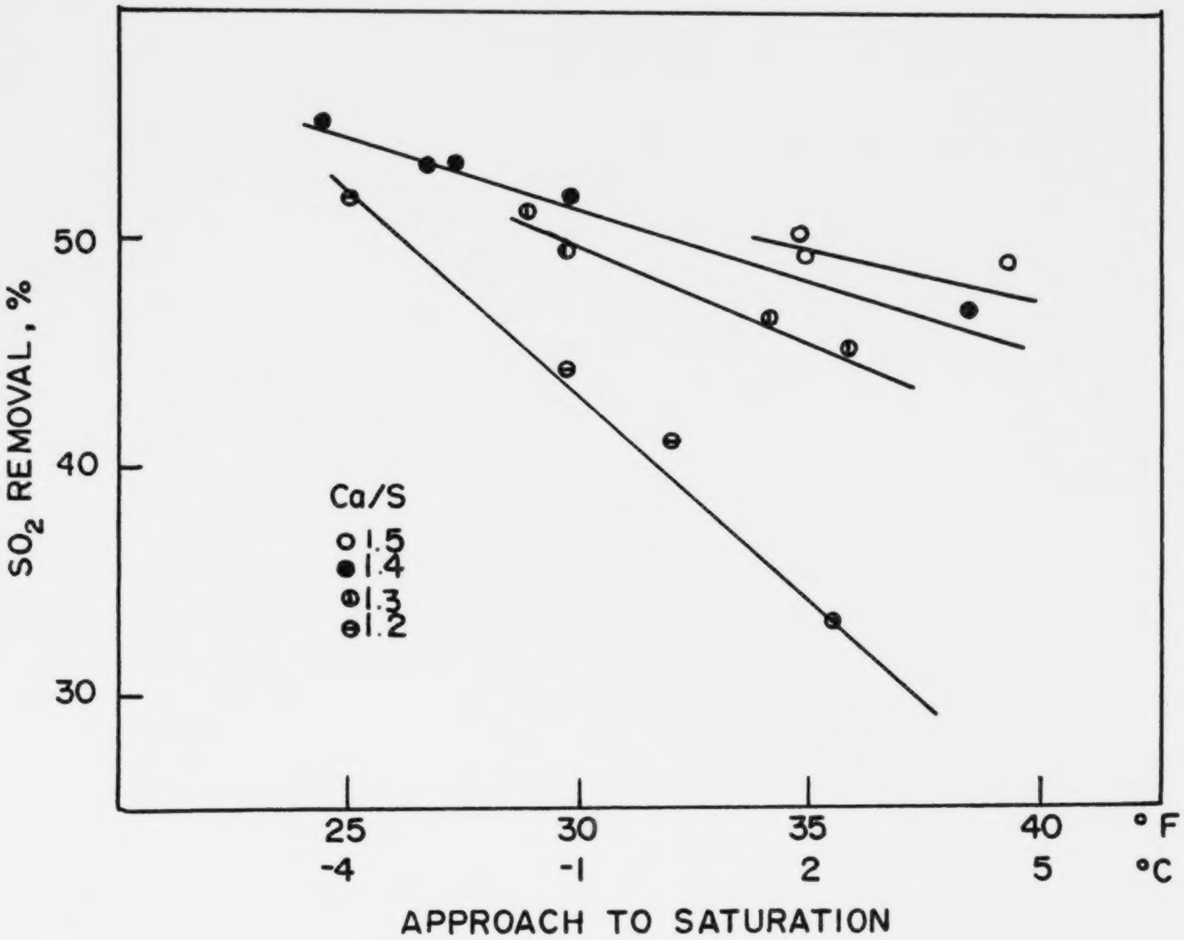


Figure 5. Effect of approach temperature on SO$_2$ removal at several stoichiometric ratios.

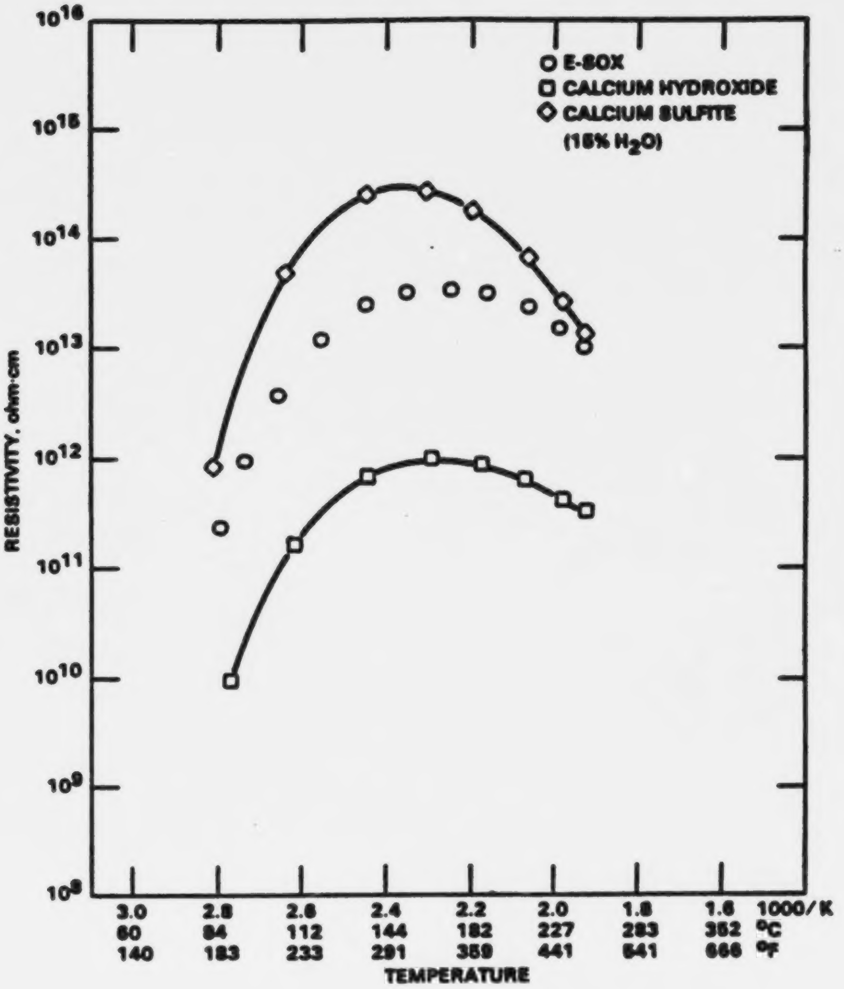


Figure 6. Resistivities of E-SO$_x$ sample compared to sample components.

TABLE 1. E-SO_x PILOT STUDIES, FIXED CONDITIONS

Air flow	1000 acfm (28 m^3/min)
Inlet air temperature	300°F (149°C)
SO_2 concentration	2000 ppm
Slurry concentration	15.5% (slaked lime)
Slurry rate	145 lb/hr (1100 g/s)
Fly ash concentration	1 gr/ft^3 (1.8 g/l)
Nozzle type	Two fluid, Casterjet
ESP configuration	Prechargers in sections 1 and 3; 1/8 in. (0.32 cm) wires in sections 2 and 4
Corona voltage	38-40 kV

FULL SCALE PRECIPITATOR EXPERIENCE FOLLOWING A LIME SLURRY FLUE GAS DESULPHURIZATION SYSTEM

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The work described in this paper was not funded by the U.S. Environmental Protection Agency. The contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

ABSTRACT

The precipitator design and test results from a large utility plant are presented and discussed.

The parameters to be considered when evaluating the capabilities of existing precipitators which may be required to follow a lime slurry FGD System retrofitted to an existing boiler system are presented.

INTRODUCTION

In the 1970's the automatic choice for particulate collection equipment following a lime slurry FGD System appeared to be a fabric filter collectors.

Although no full scale experience with precipitators on this process was available, one Utility, in 1982 decided that based on their experience with electrostatic precipitators it was their choice for the particulate collector as part of the environmental control system for their new 540 mw boiler.

SYSTEM OVERVIEW

The electrostatic precipitators are part of a turnkey Lime Slurry Flue Gas Desulphurization and Particulate Control System furnished to a utility company for a new 540 MW coal fired boiler.

The turnkey system scope included all equipment and auxiliaries from the air pre-heater outlets to the inlet of the stack, four FGD reactors, two electrostatic precipitators, lime unloading, storage and preparation system, ash handling, storage and recycle system, support steel, ductwork insulation, enclosures, piping, electrical, instrumentation and control equipment. Figure 1.

The system design criteria is given in Table 1.

The design and operation of the FGD system is beyond the scope of this papers and will be the subject of a future paper.

The particulate control system comprises two dual chamber rigid frame electrostatic precipitators operating is parallel. Each chamber having five electrical and mechanical fields in series and two cells in width, Figure 2. Each bus section is independently energized. A single pyramidal type hopper with 60^0 valley angle is located beneath each bus section. Rapping systems for the gas distribution screens, collecting electrodes and discharge electrodes are of the internal tumble hammer design. The precipitator characteristics are given in Table 2.

GAS DISTRIBUTION

As part of the system design a 1/10 scale model of the FGD and precipitator systems was built to determine gas and particulate distribution and system pressure loss. The precipitator inlet nozzles were fitted with 3 rows of perforated screens and the outlet with a single row of screens. The study was conducted at gas velocities which ensured that the flow was fully turbulent. Data measurements were taken with a vane anenometer with surveys conducted at the leading and trailing ends of the first field and at the end of the last field. The RMS deviations were 20.8 %, 13.4% and 13.5% respectively, reference Table 3A.

All gas distribution devices were installed in the precipitator in accordance with the model study. Field distribution surveys were conducted prior to start-up to verify the model study findings. Surveys taken at the trailing ends of the first and last fields were conducted with a vane anenometer with readings taken in every alternate gas passage and every 3 ft. of collectiong plate height. Initial surveys did not verify the distribution found on the model and some modifications were required at the inlet screens.

The final distribution surveys gave RMS deviations of 13.3% and 13.6% at the end of the first field and RMS deviations of 11.3% and 11.0% at the end of the last field for precipitators A and B respectively, reference Table 3B.

RESISTIVITY

Particulate samples were taken from the precipitator test sampling locations and hoppers in order to measure the resistivity. While as expected there were minor variations in resistivity between different location samples Figure 4, typically shows the resistivity for the particulate based on a 131°F dewpoint and field strength of 1 KV/CM. For comparative purposes a resistivity curve for 100% Wyoming sub-butuminous coal fly ash based on a dew point of 110°F and field strength of

1 KV/CM is also given. Wyoming sub-bituminous fuel is also burned at the plant.

It can be readily seen that with the FGD particulate at temperatures in the range of 160 - 180°F the corresponding resistivity is very low in the order of 10^7 - 10^8 ohm-cm, whereas for the Wyoming sub-bituminous fly ash at temperatures of 280°F to 320°F the resistivity is in the range of 2-5 x 10^{12} ohm-cm. This comparison again demonstrates the ease of precipitability of the FGD particulate.

ASH/PARTICULATE ANALYSIS

The fuels fired range from Wyoming low sulphur sub-bituminous coal to high sulphur bituminous coal with normally a blend of the two coals giving a sulphur in the range of 1.5 - 1.7% on a dry basis. Table 4 shows the comparison between a blended coal fly ash and the FGD particulate as seen by the precipitators. As expected the only significant difference is a 70% increase in lime resulting form the lime slurry injection in the FGD system.

PERFORMANCE TEST DATA

Performance testing of the FGD system and electrostatic precipitator was made after 6 months of system operation to determine compliance with Federal, and State Regulations and contractual guarantees. The results of the electrostatic precipitator tests are given in Table 5.

The inlet concentrations to the precipitator reflect boiler flyash and lime slurry reagent.

The precipitator removal efficiencies ranged form 99.96% to 99.97% with emission concentrations ranging form 0.0038 to 0.0047 grs/DSCF. The emissions as determined by Fd method ranged from 0.0073 to 0.0038 lb/MBTU. All emissions being less than one quarter of the 0.03 lb/MBTU permissible under Federal Regulations.

The performance tests fully confirmed the ease of ash precipitability and excellent quality of the precipitator installation.

All performance tests were conducted with the transformer rectifiers operating in the Semipulse^R mode with overall control by the EPCC energy management system. The transformer rectifier power consumption ranged from 505 to 513 KW HR/HR with corresponding specific power range of 278 to 284 watts/1000 ACFM. This power consumption represented less than 0.1% of the megawatts generated by the boiler.

Surveys conducted below the precipitator collecting plates indicated minimal if any by pass in the hopper areas.

The gas distribution met the contractual requirement and was considered acceptable for the designed performance requirements.

PRECIPITATOR OPERATION

The transformer rectifier controllers are Flakt EPIC microprocessors which include the standard Semipulse^R millisecond pulsing feature and are integrated in to a Flakt EPCC energy management system.

Initial operation of the precipitators demonstrated the particulate was readily precipitated with stack opacity levels being in the order of 1-2%. Continued operation indicated no deterioration of precipitator performance Voltage/current curves, typically as given on Figure 3., also demonstrated the precipitability characteristics of the particulate. The curves illustrate progressive increases in maximum secondary current and some decrease of secondary voltage from first to last field as expected in the precipitation of well conditioned particulates. No indication of detrimental back corona was found.

The ease of precipitabilty resulted in two conditions one offering power saving opportunity the other requiring modifications to the the ash removal system timing sequence.

An extensive test program was conducted using the Semipulse^R and energy management features of the control system to determine the optimum charge ratios for each precipitator field such that the transformer rectifier power savings were maximized without any increase in precipitator emission as indicated by the stack opacity meters.

The optimum charge ratios for the precipitators were determined to be field 1- 5:1, field 2- 1:1, field 3- 1:1, field 4- 5:1 and field 5- 7:1.

In conjunction with the charge ratio test program rapping frequency adjustments were made with particular emphasis to the last field to ensure rapping losses did not produce visible emission spikes.

The ease of particulate precipitability did however result in estimated particulate collection of an greater than 90% in the first field hoppers, somewhat higher than originally considered. The original sequence of hopper evacuation was unable to be completed prior to a full hopper alarm condition on the first field. Resequencing of the hopper evacuation system within the capability of the ash system and changes to the first field charge ratio resolved the problem to an acceptable level.

PRECIPITATOR SYSTEM INSPECTION

A recent precipitator inspection at the annual boiler outage was made to ascertain the internal condition of the precipitators after 18 months of service and to examine for any indication of corrosion.

Although no indications of insulator tracking or breakage were apparent from the operation of the precipitator some evidence of electrical tracking was apparent on some of the discharge electrode drive insulators and support insulators, 3 of which had evidence of cracks and were replaced. In addition some evidence of corrosion was found in the insulator housings as a result of condensation. Further investigation indicated that an increase in the insulator heater thermostat setting was required to eliminate any condensation.

Although all access door seals were in good condition some indications of initial corrosion were present in the casing steel adjacent to the doors. Additional thermal insulation is being installed in these areas to minimize this condition.

The precipitator internal inspection was performed with no "off line" rapping in order to get a better appreciation of both the collecting and discharge electrode conditions.

Significant deposits 1/2" thick of particulate were found in all bus section of the first field indicating that with the high particulate loadings and high flue gas moisture content the precipitated particulate inlet field had tendency to form agglomerates which were difficult to remove even with high acceleration rapping.

Particulate deposits on bus sections in the second field were no greater than 1/16" on the discharge electrodes and 3/16" on the collection electrodes.

The discharge electrodes in all subsequent fields had no build-up whatsoever and the maximum particulate layer in any area the collecting system was 1/16". The particulate layer, however, was flaky in nature and iron oxide appeared to be present. Measurements of the collecting electrode thickness, however, did not indicate any metal loss. Additional studies are being conducted to determine if corrosion of the collecing electrodes is taking place.

SUMMARY

The use of electrostatic precipitators to remove particulate following a lime slurry flue gas desulphurization system is a viable alternative to a Fabric Filter system.

No comment is offered in this paper, regarding the comparative merits of precipitators and fabric filters on the additional SO_2 removal in the particulate cleaning system. This will no doubt will be the subject of other papers.

Utilities who in the future may be required to reduce SO_2 emissions have several options to achieve this.

If a lime slurry flue gas desulphurization system is an option and the boiler system has existing electrostatic precipitators which are reasonably sized and good working order there is evidence to indicate that the Federal particulate emission level may be achievable without replacement of the existing precipitator.

PROCESS PARAMETER SUMMARY

- Lower Gas Temperature
- Reduced Gas Flow
- High Inlet Particulate Loadings
- High Removal Efficiency
- Low Resistivity Particulate
- Readily Precipitated Particulate
- Low Power Consumption Possible with Semipulse and Energy Management

PRECIPITATOR CONSIDERATIONS

- Specific Collecting Area
- Sectionalization
- Transformer Rectifier Capacity And Controls
- Internal Condition
- Gas Distribution
- Rapping Intensity - Rapping Sectionalization
- Insulator And Hopper Heating
- Thermal Insulation And Cladding

FIGURE 1

PLANT ARRANGEMENT

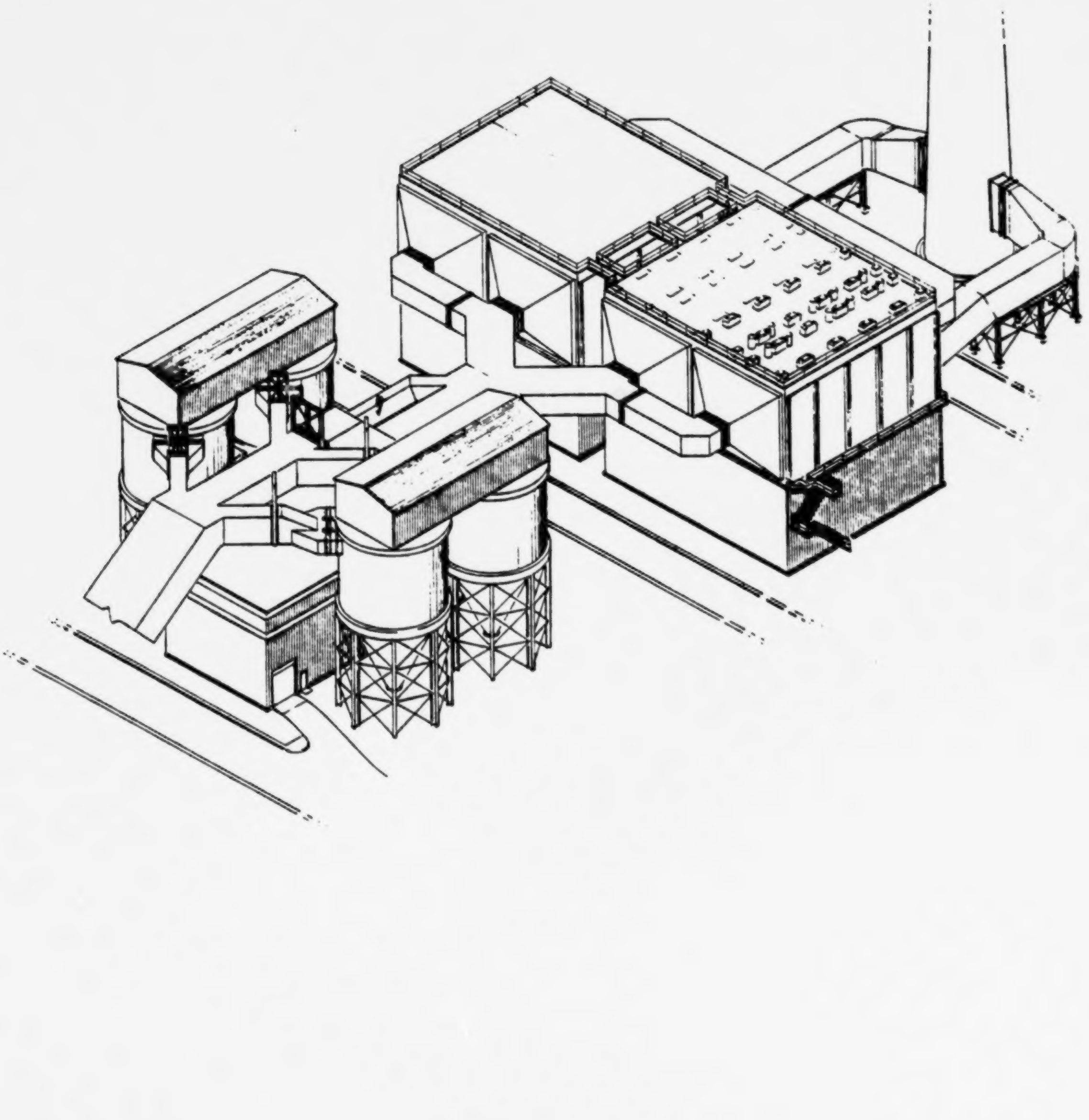

FIGURE 2

PRECIPITATOR BUS SECTION ARRANGEMENT

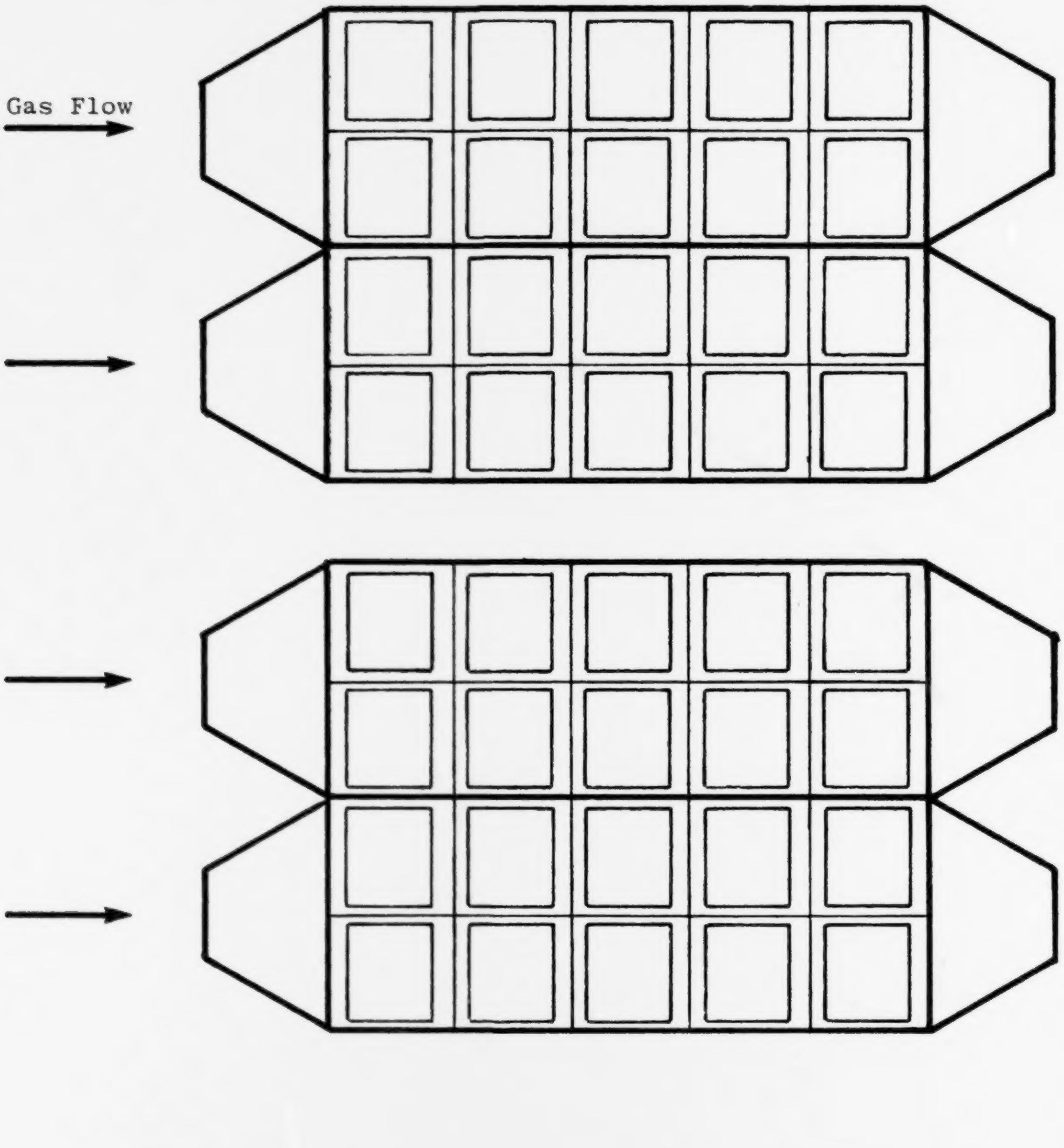

FIGURE 3

TYPICAL VOLTAGE/CURRENT CHARACTERISTICS

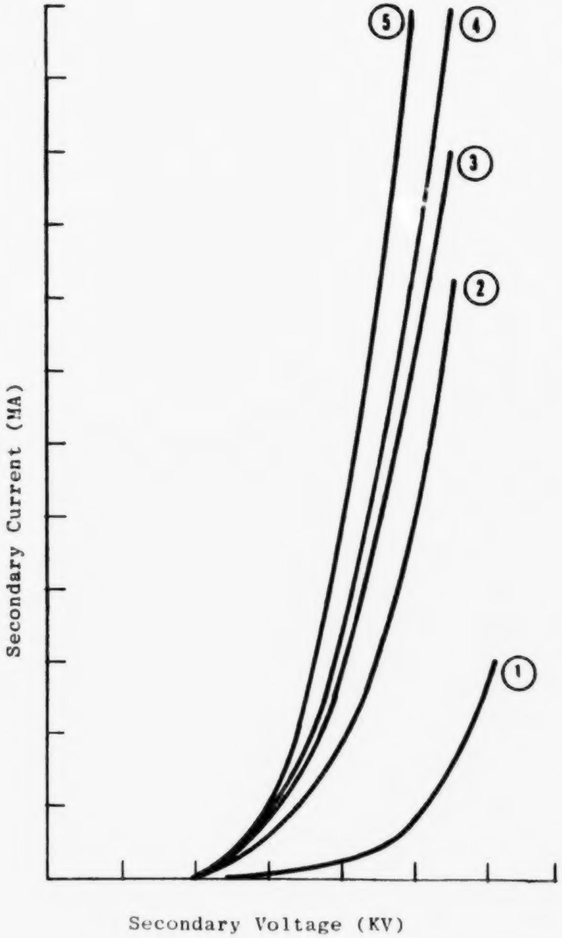

FIGURE 4

RESISTIVITY

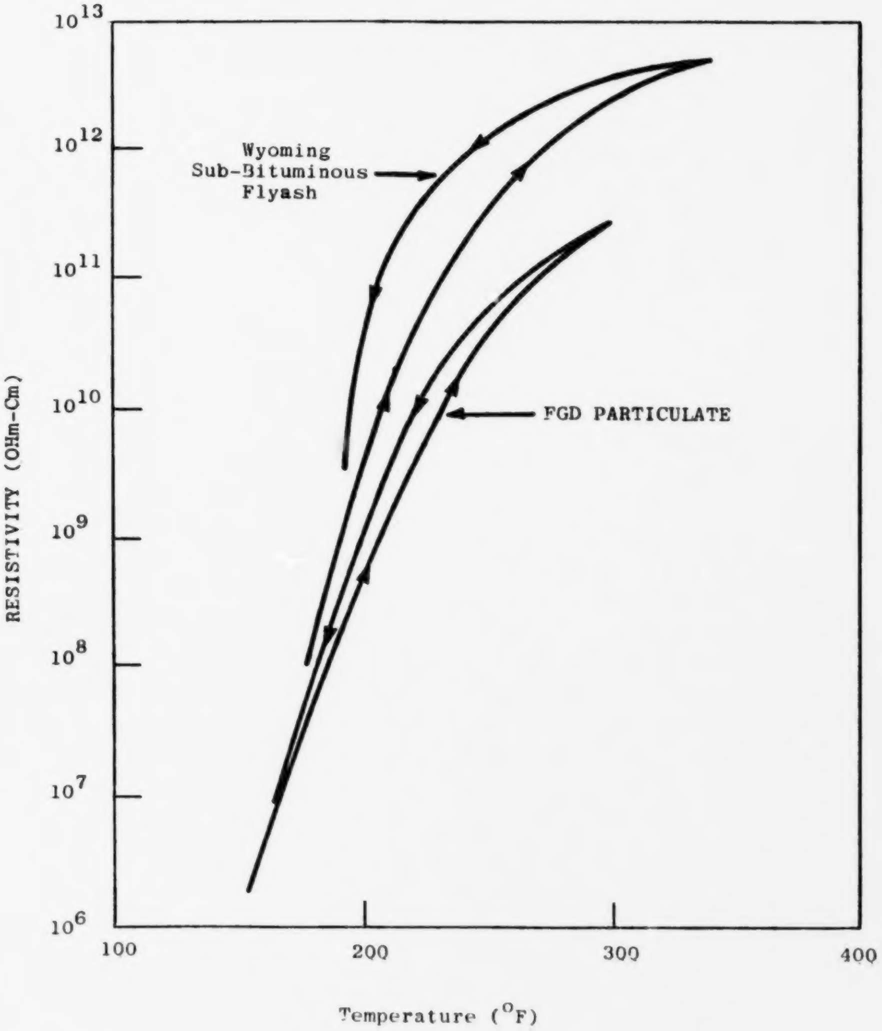

TABLE 1

SYSTEM DESIGN CRITERIA

System Gas Flow Rate	1,850,000 ACFM
System Inlet Gas Temperature	250 - 300 °F
Coal Sulphur Content Range	0.4 - 1.5%
Coal Ash Content	7%
Heating Value	8,000-9,600 BTU/lb
Sulphur Dioxide Inlet Loading	430 - 1150 ppm
Sulphur Dioxide Emission	65 - 175 ppm
Electrostatic Precipitator Efficiency	99.8%
Particulate Emission	0.0286 lb/M BTU

TABLE 2

PRECIPITATOR CHARACTERISTICS

Number of Precipitator	2
Chambers per Precipitator	2
Mechanical Fields in Series	5
Electrical Fields in Series	5
Cells per Chamber	2
Collector Plate Height	45.93 ft.
Treatment Length	61.52 ft.
Transformer Rectifier Sets per Precipitator	20

PRECIPITATOR GAS DISTRIBUTION

MODEL STUDY TABLE 3A

	± 15% of Average	±40% of Average	RMS Deviation
Inlet First Field	75%	96%	20.8%
Outlet First Field	95%	99%	13.4%
Outlet Last Field	95%	100%	13.5%

IN SITU PRECIPITATOR TABLE 3B

	1.15 Average	1.40 Average	RMS Average
'A' Precipitator			
Outlet First Field	89%	100%	13.3%
Outlet Last Field	96%	100%	11.3%
'B' Precipitator			
Outlet First Field	88%	100%	13.6%
Outlet Last Field	97%	100%	11.0%

TABLE 4

		FGD PARTICULATE	FGD PARTICULATE	COAL	COAL
% Moisture, as received		0.84	0.93	14.3	14.6
% Ash, dry basis				9.6	9.8
% Loss of ignition, dry basis		3.05	3.11		
Sulphur forms, dry basis					
% Sulphur Pyritic				0.61	0.56
% Sulphur Sulphate				0.10	0.08
% Sulphur Organic				1.00	1.00
% Sulphur Total				**1.71**	**1.64**
Mineral analysis, ignited basis					
Phosphorus pentoxide %	P_2O_5	0.66	0.58	0.62	0.64
Silica	SiO_2	23.5	21.4	27.7	27.7
Ferric Oxide	Fe_2O_3	7.5	7.9	12.0	12.0
Alumina	Al_2O_3	11.4	10.2	11.7	11.8
Titania	TiO_2	0.74	0.65	0.74	0.75
Lime	CaO	32.6	33.2	18.9	19.0
Magnesia	MgO	3.23	2.93	2.92	2.90
Sulphur Trioxide	SO_3	19.0	21.0	24.5	24.0
Potassium Oxide	K_2O	0.46	0.47	0.62	0.60
Sodium Oxide	Na_2O	0.96	0.77	0.88	0.91
Lithium Oxide	Li_2O	0.02	0.02	0.01	0.01

TABLE 5

ELECTROSTATIC PRECIPITATOR PERFORMANCE

	Test No. 1	Test No. 2	Test No. 3
Gas Flow Rate (ACFM)	1,780,000	1,828,000	1,829,000
Inlet Gas Temperature (°F)	159	156	156
Inlet Particulate Concentration (gr/DSCF)	10.57	10.45	11.66
Outlet Particulate Concentration (gr/DSCF)	0.0040	0.0047	0.0038
(lb/MBTU) *	0.0078	0.0098	0.0073
Precipitator Removal Efficiency (%)	99.96	99.96	99.97
Opacity (%)	1.5	1.5	1.5
Transformer Rectifier ** Power Consumption (KWHR/HR)	505	509	513
Specific Power Consumption (Watts/1000 ACFM)	284	278	281

*** By the Fd Factor Method**

**** Semipulse /Energy Management**

VIDEO BASED IN-SITU DROPLET CONCENTRATION AND SIZE ANALYZER

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ABSTRACT

Data concerning the droplet size distribution of liquid sprays is often needed in research and diagnostic explorations of scrubbers and mist eliminators. In addition, information concerning the spatial and temporal variability of droplet concentrations and size distribution is often desired. The methods that have been available in the past for making such measurements have been generally less than satisfactory because of problems relating to various combinations of time resolution, traverse capability, representative sampling, sizing range, size resolution, survivability of equipment in the scrubber environment, and cost. A video based droplet size and concentration measurement system has been constructed which provides a solution to most of the above problems. A description of the droplet analyzer and how it functions is provided together with exemplary data from field applications.

INTRODUCTION

One of the most difficult aspects of scrubber research and problem solving over the years has been associated with obtaining data on spatial and temporal distributions of droplet size and concentration. The sizes of the droplets in a scrubber play an important role in the performance (efficiency) of the scrubber collecting either particles or gases. The spatial distribution of the liquid similarly plays an important role. These factors, together with the gas velocity distribution, not only are major factors in determining the collection efficiency provided by a scrubber, but also in mist eliminator (ME) performance (and malperformance). Thus knowledge of spatial and temporal variations in droplet concentrations and size distributions is potentially of great value in diagnosing the causes of problems in scrubber/mist eliminator operation which in turn are needed if solutions to the problems are to found in the most expeditious way.

Several methods have been applied in the past, with varying degrees of success, for measuring or estimating droplet size distributions and concentrations. Methods for size measurement have included droplet impingement on coated slides or sensitized paper (1,2), hot wire devices (3), high speed photomicroscopy (4,5), and the use of inertial separators such as impactors or cyclones(6). Each of these has operational limits which restrict its utility. The coated slide/sensitive paper techniques are difficult to apply in high concentration streams and typically provide single point "snapshots" of the size distribution. Sampling and analysis techniques for these two methods are too labor intensive to make wide area traverses for spatial distribution or long term measurements for time resolution feasible. Inertial separators provide data only for sizes smaller than about 15 micrometers while most of the droplet spectrum of interest is in substantially larger sizes. Further, it is difficult or impossible to obtain time resolved data using the inertial devices. The photographic technique offers the possibility of spatial and time resolution but unless applied with motion picture technology has a low data rate; it also has a long analysis time, even if automated image analyzers are used to examine the resulting photomicrographs.

Automated video based droplet analyzers have been used for at least 20 years; some with online real time data analysis (7) and others with much slower offline analysis (8). Because of equipment size these techniques have generally been limited to laboratory application. However, with the advent of small light-weight solid state video cameras it became practical to construct a real time video based droplet analyzer for field application. Such a system was expected to be of great use in diagnosing problems related to mist eliminator carryover in scrubbers at electric utility power stations. Hence the Electric Power Research Institute contracted with Southern Research Institute through the Stearns-Roger Division of United Engineers and Constructors to construct such a system and provide data from it for several sites which were experiencing ME problems. This paper provides a description of the system which was constructed and some results typical of the data that has been obtained using it.

DESCRIPTION OF THE VIDEO DROPLET ANALYZER (VDA)

The droplet analyzer system consists of a sectionalized probe which provides a working length of up to 6 m (20 feet), a probe operator's console and monitor, analyzer console, and a microcomputer for data collection, storage, and analysis. The probe is designed to work through a 10 cm (four inch) or larger access port. The principal components of the system are shown in Figure 1. The probe is connected to the analyzer console by 75 m (250 feet) of cable permitting the analyzer and computer to be located well away from the sampling area.

The probe carries an illuminator, a small solid-state video camera, and its associated "macro" lens. A choice of lenses and extension tubes are available to provide particle sizing ranges from 3-450 micrometers to 45-6500 micrometers diameter with corresponding active fields of view ranging from 0.75mm x 1.0mm to 12mm x 15mm. The size resolution at any magnification is equal to the minimum measurable size for the magnification. Data are obtained in 220 equal interval size channels, although usually only the first 150 channels are retained. Back lighting is used to produce dark droplet images on a bright background. The illuminator and camera unit is shown partially disassembled in Figure 2. The illuminator is suspended from and held directly opposite the camera housing on four small tubes. The view volume is located in the center of the large open area between the camera and illuminator housings. Illumination is provided by a high intensity strobe lamp with a flash duration of 250 nanoseconds. The flash lamp is synchronized to fire during alternate vertical retrace periods of the video camera. The short flash duration effectively eliminates blur from droplet motion. The illuminator and camera lenses are protected from fouling due to entrained liquids and solids by baffled shrouds and a purge air system. The purge air is supplied via a hose from a small compressor located adjacent to the operator's console.

Data acquisition and preliminary analysis take place in real time at a rate of 30 data frames per second. Objects in the field of view are detected by analysis of the video signal level during each scan line. On detection of an object, a timing loop is started which results in the analyzer's looking for a recurrence of an object in a similar horizontal location on succeeding scan lines until it is no longer found. During this process the number of scan lines on which the object was detected is counted. The vertical height of the object is directly proportional to the number of scan lines which it intercepts. Thus the intercepted line count is a direct measure of its height (or diameter if the object is spherical as is the case with small droplets). Focus information is obtained from the shape of the video signal of the image. When a particular image is no longer detected, its intercepted line count and focus information are transmitted to a microcomputer for processing. Four parallel counting (timing loop) circuits are incorporated in the analyzer, permitting simultaneous analysis of up to four particles whose images fall on a common video scan line. After the data for an object has been transmitted to the computer the counting circuit which handled the object is cleared and freed to handle another object. Thus a large number of droplets can be measured in a single video frame. Output of data from the image analyzer to the computer is suppressed if any part of the image falls outside the active field of view. This eliminates incorrectly sizing any droplet whose entire image is not available for measurement but makes the effective field of view dependent on the drop size. Major details of the image analysis are depicted in Figure 3.

Upon receiving the image height and focus data from the analyzer the computer clears the analyzer timing loop which generated the data, freeing it for another image, determines whether or not predetermined focus criteria were met, and if so increments the number counted in the appropriate size bin. Partial data are updated once each second on the computer's video display and full data in tabular and graphical forms are available at the end of an operator selected sampling period which can range from 1 to 255 seconds. The data are also stored on floppy disk for further analysis at a later time.

Video monitors in the probe operator's console and at the image analyzer/computer location provide direct views of the camera's video output and diagnostic information on the performance of the image analyzer circuitry. The video signal from the camera can also be recorded for later inspection and analysis.

The dimensions of the active sensing volume are determined by the height and width of the field of view and the depth of field. The field height and width are fixed by the geometries of the lens and camera. However, the depth of field varies with illumination intensity and particle diameter and must be measured empirically. The dimensions of the sensing zone are small enough for all usable magnifications that, at the velocities encountered in scrubbers and ducts, the sensing zone is swept clear and a new sample is introduced between flashes. The total gas volume sampled at a point can thus be calculated from the sensing zone dimensions and the number of video scans (flashes) which were analyzed. Assuming spherical droplets, their volumes can be calculated from their measured heights (diameters). These volumes can then be summed to obtain the total volume of the particles measured, thus permitting the liquid loading (concentration) in the size range spanned by the instrument to be calculated. Assuming that no significant contribution to the liquid loading is due to droplets having diameters outside the instrument range, the calculated loading can be taken to be the actual liquid concentration in the gas stream. The shapes of plots of measured liquid volumes by size channel can be used to access the validity of the assumption that no significant droplet volume fell outside the measurement range in any given data set.

ANALYZER CALIBRATION AND PERFORMANCE VERIFICATION

Primary calibration of the system is based on glass beads of known sizes mounted on microscope slides. These beads are placed in the view volume and the line count is noted for calibration of diameter versus number of intercepted lines. The slide is then traversed through the depth of the sensing zone by means of a micrometer stage to determine the depth of focus for the size in question. The acceptable focus criteria are adjusted so that the image is considered out of focus if the indicated size differs from the true size by more than ten percent or one line count, whichever results in the greater relative error (ie. one line count for small objects and ten percent in size for large objects). This primary calibration need only be done once in the laboratory for each lens/illuminator setup. Thereafter, depth of field checks at one or two sizes suffice for obtaining corrections for any immediate application.

The measured size spectrum for a mixture of two sizes of glass beads, 450 to 550 μm and 900 to 1250 μm diameter, is shown in Figure 3. The mixture was made up to be approximately 50/50 on a weight basis. This mixture was slowly poured through the sensing zone of the analyzer as part of the verification of the system performance.

Further verification of the performance of the system was obtained by measuring droplet transport rates in the laboratory both directly and by means of the VDA. Two such trials were performed on aerosols which had mass median diameters of about 100 micrometers. The VDA was configured for measurement in the range from 3 to 450 micrometers for these tests. The VDA results showed 80% to 94% of the liquid transport rate measured by direct collection of the liquid that passed through the sensing zone. Comparisons have also been made of the size distributions of liquid sprays from nozzle types for which data were available from other techniques - Malvern optical particle size analyzers and KLD hot wire devices. In each case, the results from the VDA were similar to those from the other technique in the size ranges which they had in common.

CALCULATION OF LIQUID TRANSPORT RATES

If the gas flow is horizontal at the measurement plane and the gas velocity is known, the system also provides a direct measurement of droplet flux (or liquid transport rate). In this case, the velocities of the droplets in the direction of the gas flow can be taken to be equal to the gas velocity. However, if the gas flow is vertical the situation becomes more complex. In the latter case, the droplet flux (in the size range measured) is given by the equation:

$$\text{Flux} = \int_{D_{min}}^{D_{max}} c(D) \times (V_{gas} - V_s(D))\,dD$$

where D_{min} = minimum measurement diameter

D_{max} = maximum measurement diameter

$c(D)$ = concentration of drops having diameter, D

V_{gas} = local gas velocity (positive for flow upward)

and $V_s(D)$ = settling velocity for drops of diameter, D.

For vertical gas flow in the upward direction, small droplets will be moving upward at approximately the velocity of the gas. As the drop size increases the upward velocity of the drops will decrease until the size of the droplets becomes large enough that the settling velocity equals the gas velocity. Droplets having this size, defined as the limiting droplet diameter, will remain virtually suspended until they either decrease in diameter from evaporation or grow by agglomeration. Droplets having diameters greater than the limiting diameter will fall back against the gas flow. The instantaneous values of the limiting diameters will vary with time because of moment to moment velocity fluctuations but values based on local average velocities should be representative. The calculation of liquid fluxes for the case of vertical gas flow thus requires information on the droplet concentration by size, the local gas velocity at each sampling location, and the settling velocities of the droplets by droplet size. The droplet settling velocities can be calculated but the local gas velocities must be separately measured.

Traverses of a sampling plane can be made in two ways. In the first, the probe is inserted through conventional 10 cm (four inch or larger) sampling ports and data is obtained at points along horizontal lines whose spacing is set by the port locations. Because mist eliminator entrainment problems may be associated with highly localized areas across their faces, it was deemed desirable to have the ability to perform a continuous scan over large areas in order to locate trouble spots. Consequently a swiveling probe mount was constructed to provide such a capability. The swiveling mount permits the probe to be swung through an arc slightly greater than 90 degrees while moving in or out over its full 20 foot length. Thus a continuous survey of an area having a 20 foot radius and a slightly more than 90 degree included angle can be obtained. Potentiometers installed in the mount provide information on the position of the particle sensing zone to the computer; this data is logged with the droplet data. This permits accurate repositioning of the probe to obtain more detailed data after identifying areas of special interest. Operation with the swiveling mount does require a large amount of free space around the sampling area and an access port much larger than is commonly used (12 inch or greater in diameter in order to use the full angular swing capability of the mount). The probe is sectionalized in lengths of about 60 cm (22 inches) to minimize free space requirements and access problems in the sampling area. Because of problems in keeping the optical surfaces clean and dry, the probe must be kept horizontal during operation.

By removing the illuminator from the probe, replacing it with a flood light type illumination system, and replacing the "macro" lens with a conventional lens, the probe can be used to examine interior surfaces, etc. within the duct while the scrubber is in operation.

FIELD RESULTS

To date the system has been used at five sites, making measurements around mist eliminators on seven scrubbers and characterizing the spatial distribution of droplets under a variety of conditions in a spray dryer application. It has also been used to measure the droplet size distribution produced by several nozzles being used for spray dry applications.

Figure 5 shows a typical data set taken at the inlet to a mist eliminator on a scrubber installation at a large utility pulverized coal boiler. Measured droplet concentrations by size are shown on the left in the figure. The volumes are in terms of the volume of the droplets in liters per cubic meter found in each of the measured size intervals. In this instance, the droplet size distribution appears to be multimodal with much of the volume in droplets in the one to three millimeter size range. Even larger droplets were present but not shown in the limited size span of the illustration. At the time these data were taken the gas velocity in the sampling zone was such that the limiting droplet diameter was about 1300 micrometers. That is, droplets smaller than 1300 micrometers (1.3 mm) would be carried upward by the gas flow while those larger than this were either falling back against the gas flow or would eventually do so. The liquid transport rate at the measurement plane is shown on the right side of the figure. The transport rate is given as cubic centimeters of droplets per second per square meter of tower cross-section. As droplet sizes increase the net upward velocity of the droplets decreases with small droplets moving at the gas velocity while those near the limiting diameter have near zero velocity. Thus although in terms of concentration, the distribution is dominated by droplets larger than 1000 micrometers, the actual carryover to the mist eliminator is dominated by droplets in the 200 to 900 micrometer size range. If there were no sources injecting large droplets upward into the gas stream below the mist eliminator and no sources above the mist eliminator, the total negative flux, or fallback, should equal the total positive flux, or carryover, if the mist eliminator operated at 100% efficiency.

Figures 6 and 7 show results obtained at the outlet of a ME under two load conditions (40% of full load and full load). As would be expected, the carryover rate was much larger at high load than at low load. The liquid influx to the ME was lower at low load and the mechanisms that cause entrainment from the ME are not as effective at the lower velocities that result from the reduced gas flow. At the high load condition the total carryover from this module was about two liters per minute (0.5 gallons per minute). A companion module to the one from which the data in Figures 6 and 7 were taken had a measured total carryover rate of 12 liters per minute (3 gallons per minute) when the unit was operating at 85% of full load. In terms of detailed spatial distributions, the carryover rates varied by more than a factor of 100 over the zones mapped in each of these two scrubbers and the zones with high carryover rates tended to zones of much greater than average superficial velocity. Thus improvements in the uniformity of flow distribution might be expected to result in reduced carryover rates. In addition, inspection of the ME structure in regions of high carryover can reveal mechanical causes for high entrainment rates in some instances.

In some instances temporal variations can reveal important information as illustrated in Figure 8. This figure shows the variation in concentration at a single location above the exit of a ME through three ME underspray wash cycles. As can be seen, the droplet concentrations increased dramatically when one specific zone of the ME was being washed. That zone in this instance was located approximately 8 m (25 feet) from the measurement location which was itself less than 50 cm (2 feet) above the ME face. Because the area producing the carryover was located so far from the measurement point, only a small part of the actual carryover would have been seen. It was in fact surprising that any was seen, and the true carryover must have been many fold greater than that measured at the selected location. Conservative estimates of the average droplet carryover from this and another similar scrubber tower were of the order of 150 liters per minute (40 gallons per minute) each.

CONCLUSIONS

The Video Droplet Analyser has proven to be a field usable system capable of providing data that is virtually unobtainable in any other fashion regarding droplet size distributions, concentrations, and transport rates. Useful information has been obtained to date for purposes of mist eliminator problem diagnostics and in quantifying processes in spray dry SO_2 control applications. Other applications might include fundamental research on scrubber performance, verification of design predictions for scrubber spray distributions, and verification of mist eliminator performance.

ACKNOWLEDGMENTS

The work described in this paper was done under the sponsorship of the Electric Power Research Institute as part of an ongoing program of Mist Eliminator Field Studies. The prime contractor for the work was the Stearns-Roger Division of United Engineers & Constructors, Inc. under EPRI contract RP 2250-3. The EPRI project manager was Mr. Richard G. Rhudy and the Stearns-Roger project manager was Mr. Robert J. Keeth.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

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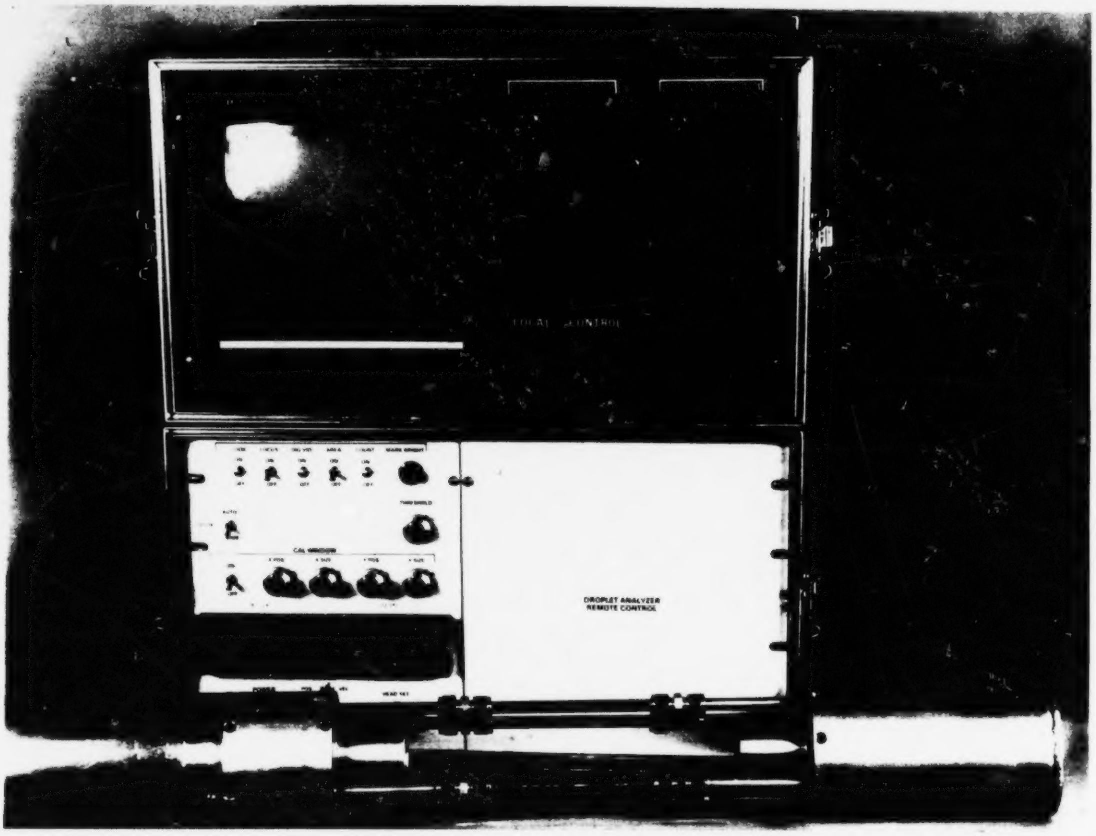

Figure 1. Video droplet analyzer electronics and probe end.

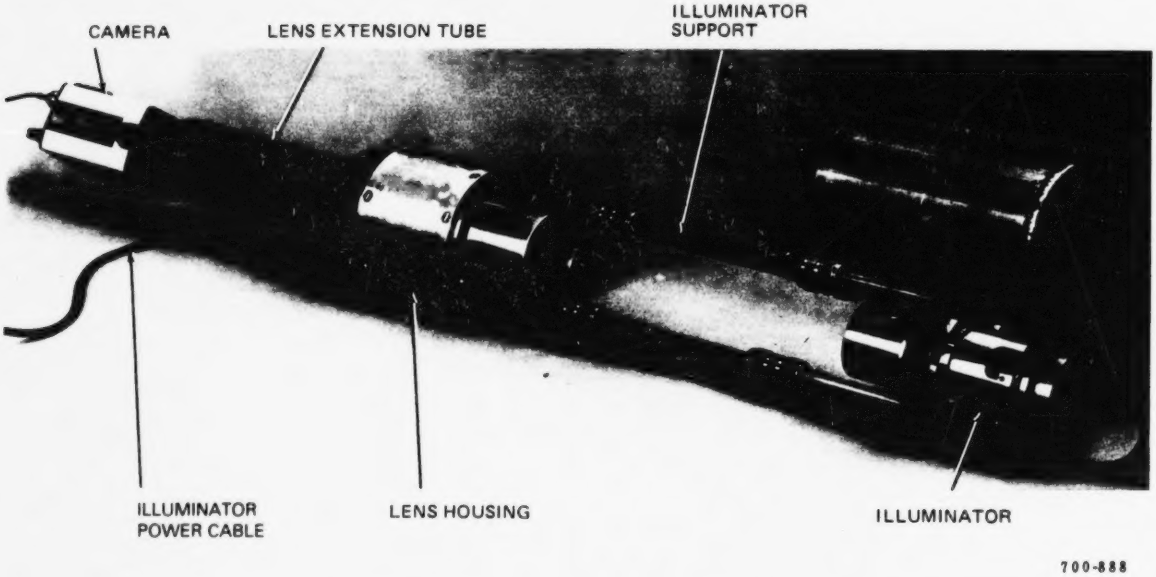

Figure 2. Video Droplet Analyzer Probe end partially disassembled showing camera, lens extension, illuminator and illuminator power cable.

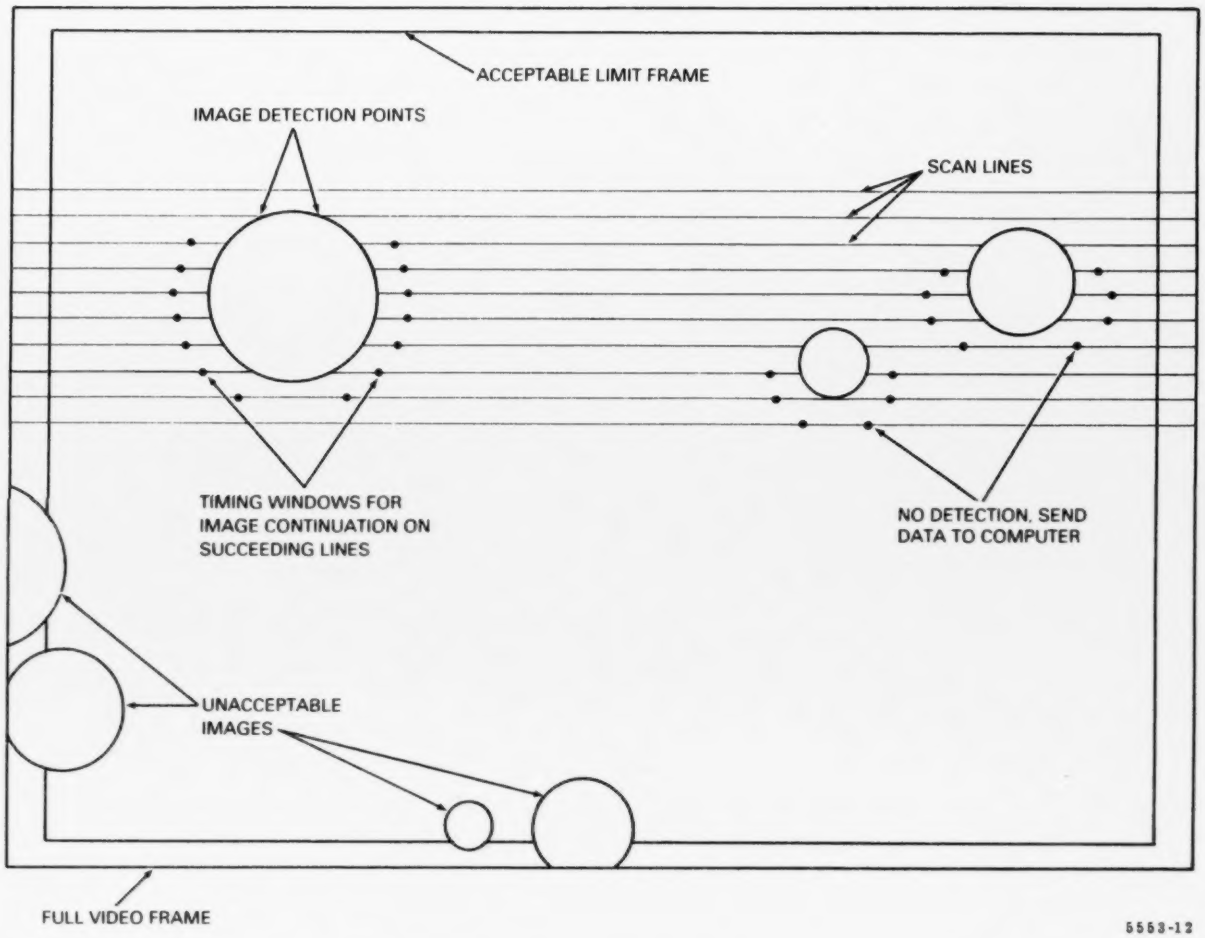

Figure 3. Depiction of image analysis details for the SRI Video Droplet Analyzer.

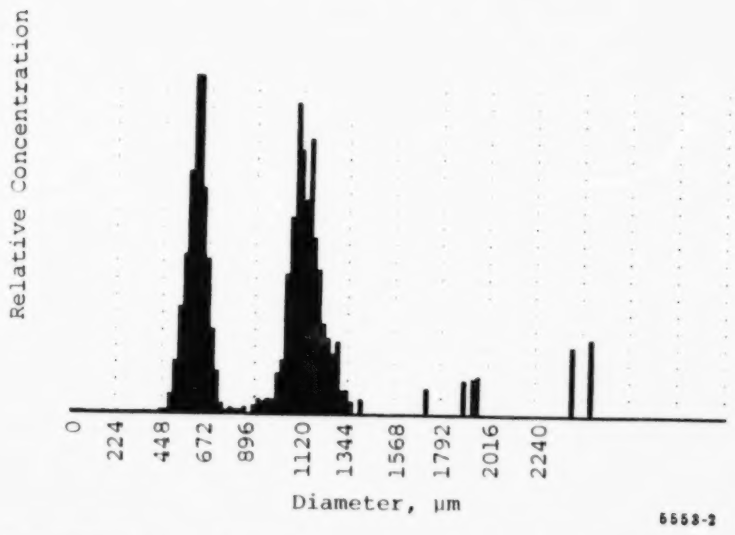

Figure 4. *Size spectrum of a mixture of glass beads of two size classes, 450 to 550 µm and 900 to 1250 µm, as measured with the Video Droplet Analyzer.*

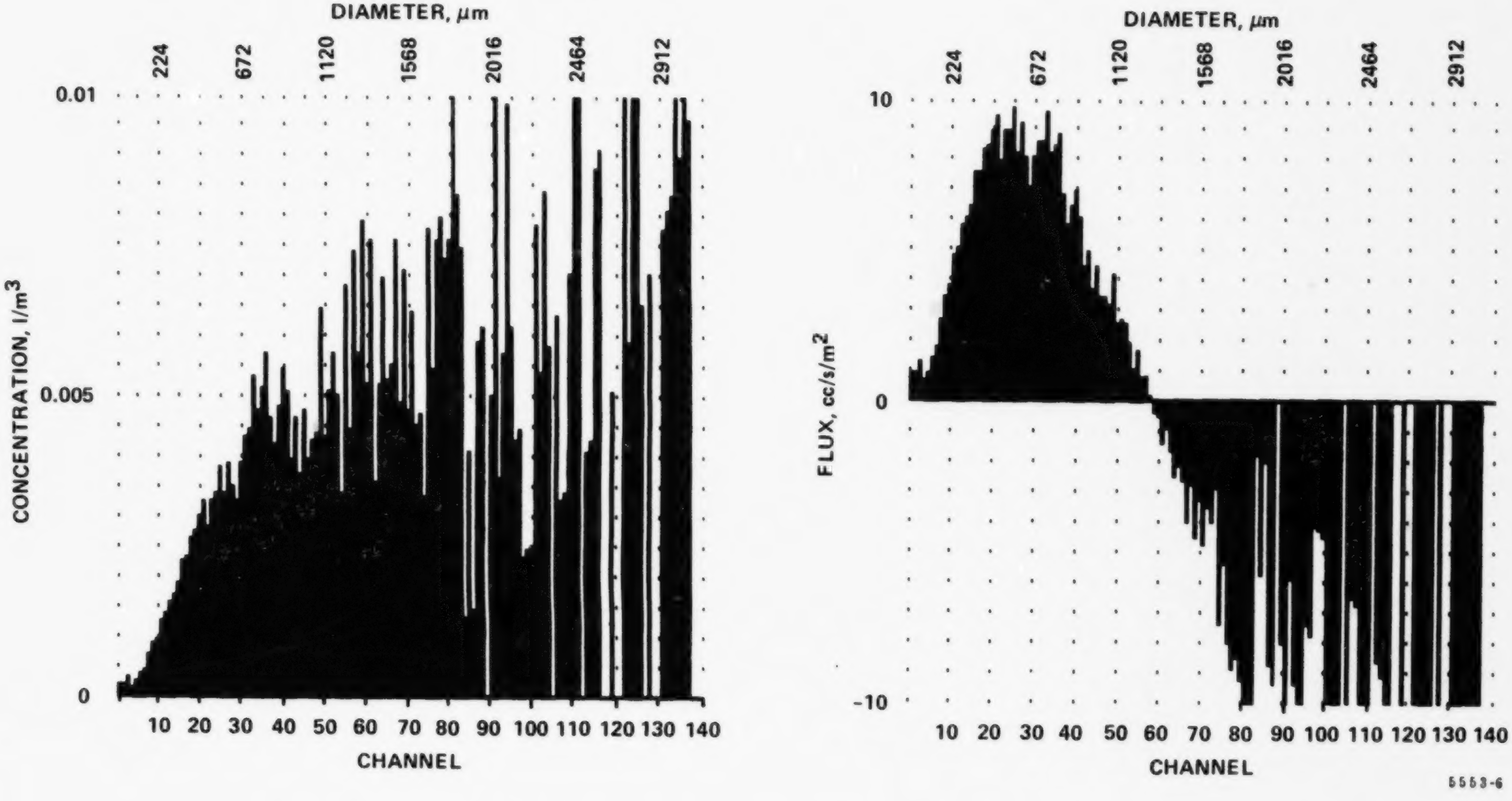

Figure 5. Concentrations and liquid fluxes by channel for a twelve minute sampling period at a ME inlet. Channel width is 22.4 μm. Positive fluxes represent carryover and negative fluxes represent fallback.

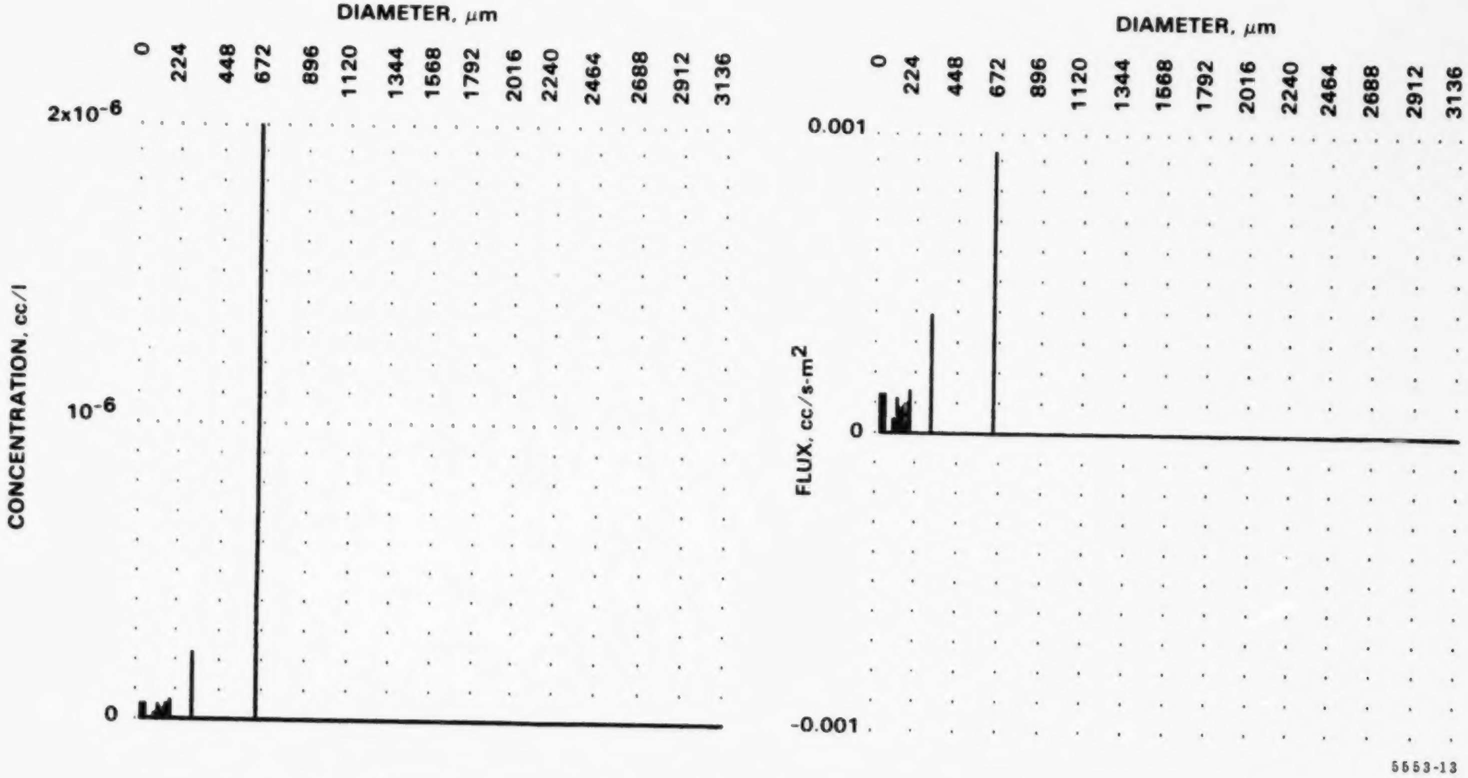

Figure 6. VDA results on a channel by channel basis for the carryover from a scrubber module outlet at a low load test condition.

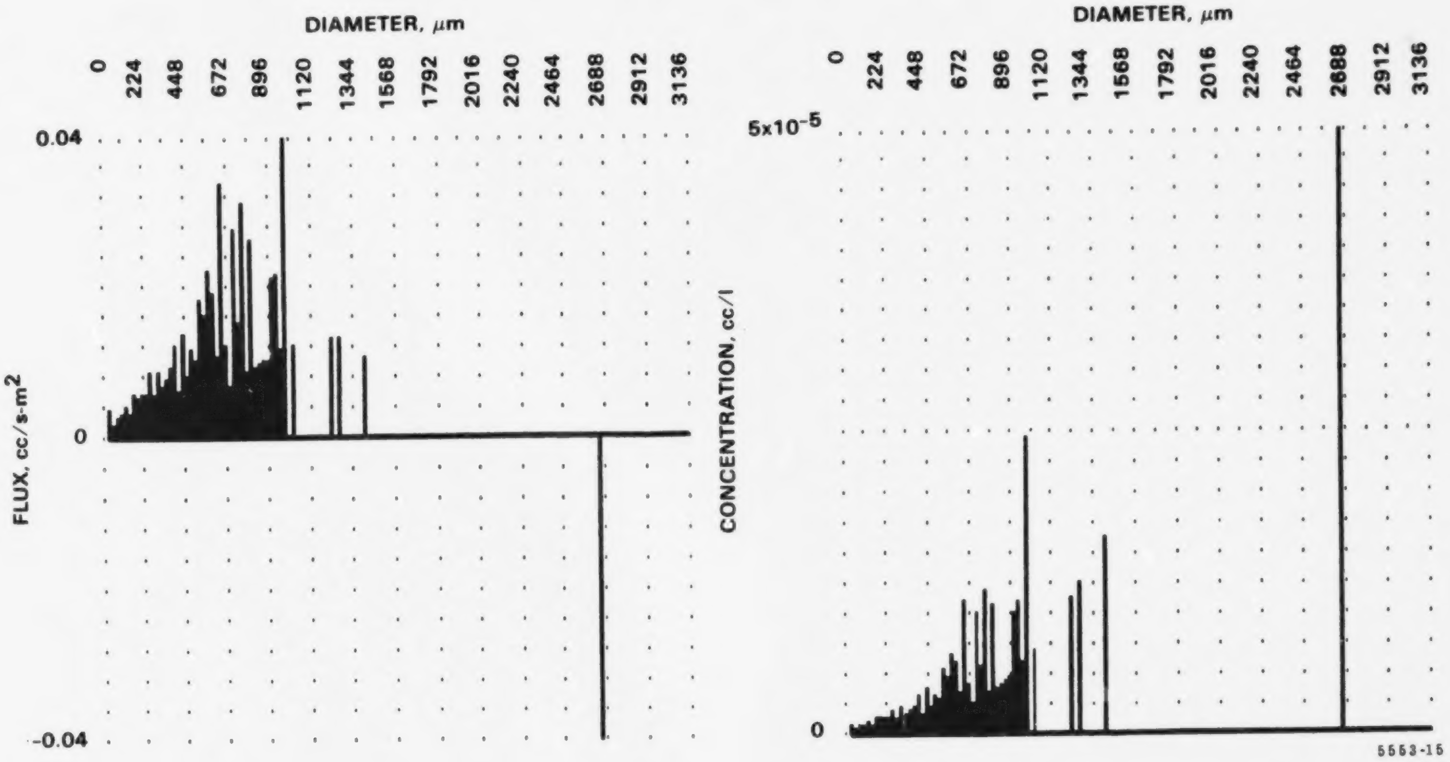

Figure 7. VDA data on a channel by channel basis for the ME outlet of the same module as in Figure 6 at the high load test condition.

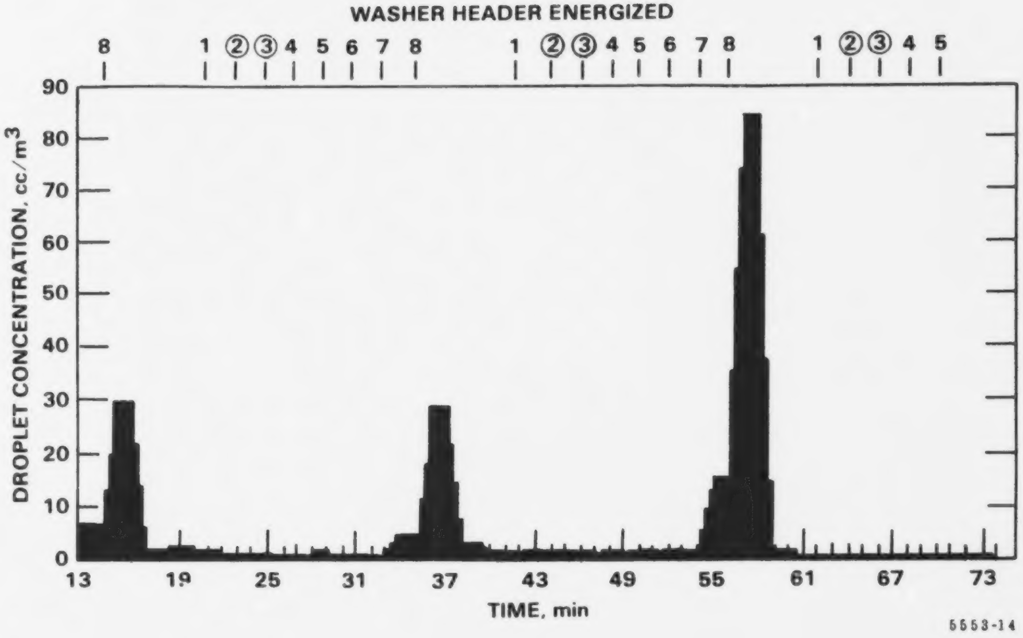

Figure 8. *Variation of droplet concentration with time at a single location above a mist eliminator outlet. Timings of ME washes are shown by wash header. The sampling location was above an area washed by headers 2 and 3.*

FABRICS EVALUATION FOR UTILITY BAGHOUSES ON COAL-FIRED BOILERS

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ABSTRACT

It is generally accepted that baghouse performance (i.e., pressure drop, particulate collection efficiency, and bag life) will be improved if optimum fabric and bag constructions are developed specifically for utility baghouse applications. Despite this common view there is no well-defined method for determining the optimum fabric/bag configuration for a specific application. The selection of bags has been typically based upon recommendations of manufacturers, operator experiences, or evaluations of trial bags for limited periods. In recognition of these factors, a new experimental method was developed for quantitatively evaluating the relative performance of different types of bags. Over the past two years, this method has been applied to study the effects of fabric design parameters on baghouse performance. This paper describes this methodology and the experimental conditions for the test program. The test results are presented in a companion paper, "Parametric Testing of Filtration Fabrics for Low-Sulfur-Western Coal Fly Ash" by Felix et al.

BACKGROUND

The Electric Power Research Institute (EPRI) fabric filter research program began in 1975. Since its initiation the program has advanced the technical understanding of fabric filter technology and contributed to the acceptance of baghouses by the U. S. utility industry. For the first several years the research was focused on characterizing the effects of baghouse and cleaning system design/operating parameters upon performance (i.e., pressure drop and particulate collection efficiency). Since the late 1970s the majority of the work was conducted under well-controlled conditions in pilot plants utilizing full-scale bags. Due to a lack of performance data for filter bags during this period, commercially available bags were chosen for the pilot tests of a type and in a manner similar to that for full-scale baghouses.

By 1984, it had become apparent to those involved with the EPRI program that the effects of fabrics and bags on baghouse performance was the single most important issue that had not been adequately addressed. Accordingly, since 1985 the program has placed increased emphasis on fabrics research. The results have been multifold: Preparation of a handbook describing conventional fabrics and filterbag manufacturing practices for utility baghouse installations (2), implementation of a compartment-by-compartment pilot evaluation of the effects of fabric texturization on baghouse performance (3), formation of an utility advisory committee to help prioritize EPRI's activities, organization and implementation of a manufacturers workshop devoted to fabric related technical issues, and the design and implementation of fabric screening studies which are the subject of this paper.

MANUFACTURERS ADVISORY GROUP

To address the issues involving proprietary properties of fabrics, to help secure test bags of a prescribed construction, and to promote the utilization of the experimental results, an advisory group was formed to assist EPRI and its contractors in the development and implementation of this program. The group is called the Fabrics and Filterbag Manufacturers Advisory Group and is comprised of representatives from the following companies:

BHA Group, Inc.
Burlington Glass Fabrics Company
E. I. du Pont de Nemours & Company (Inc.)
J. P. Stevens & Company, Inc.
Menardi-Criswell
Midwesco, Filter Media Resources Division
Owens-Corning Fiberglas Corporation
PPG Industries, Inc.
Tex Tech Industries, Inc.
Uniglas Industries (Div. of United Merchants & Manufacturers,Inc.)
W. L. Gore & Associates, Inc.

HISTORICAL PROBLEMS CONDUCTING FABRIC TESTING

In the past, fabric testing was not conducted in a scientific manner due to a combination of technical and practical limitations. These considerations are briefly reviewed below.

Limitations of Full-Scale Compartment Trials

One of the most often attempted approaches for qualifying new fabrics or bags was to install a set of "trial" bags in one or more compartments of a full-scale baghouse and to monitor its relative performance. The major problems with this approach are that it is practically impossible to conduct a well-controlled parametric evaluation (gas flow, dust loading, temperature, dewpoint excursions, etc.), the results are very site specific (i.e., unique to the fly ash and baghouse cleaning system parameters), and direct comparisons with bags in other compartments are misleading because of the compensating operating character of baghouses with respect to pressure drop. Hence, this type of evaluation is usually only useful for screening bags that may prematurely fail in a catastrophic manner.

Flow Compensating Operating Character of Baghouses

Baghouse compartments are operated in parallel. Consequently, the uniformity of flue gas flow through a baghouse depends on the resistance to flow in each compartment. Components of flow resistance include contributions from the ductwork and valves, and across the fabric and dustcake composite. The larger component is typically the resistance across the fabric and dustcake, although in some baghouses ductwork contributions can be responsible for up to 30% of total flow resistance. Those compartments (or bags) that have a relatively light dustcake, which would be the case for a compartment with new bags, will tend to have lower resistance, and consequently will take proportionally more flue gas flow. As a result the dustcake will grow at a rate faster than for bags in other compartments, and over time the flow resistance will tend to equilibrate throughout the baghouse. Hence, comparative flow, pressure drop, or dustcake buildup evaluations over a fixed period of time can be very misleading if the effects of non-uniform flow are not taken into account. It would, therefore, be more meaningful to perform comparative evaluations based on the cumulative amount of flue gas that was filtered rather than a prescribed period of time. Unfortunately, monitoring the respective compartment flows on a full-scale baghouse is generally impractical.

Duration of the Experiment

Comparative evaluations require thousands of operating hours (months) to produce clear-cut results. This is due to a variety of factors. These factors include the need to separate influences of day-to-day boiler operations and upset; the slow, continuous process of dustcake growth, which takes several thousand operating hours to achieve a state of quasi-equilibrium; and the need to include a full range of operating experiences (i.e., startups, shutdowns, and multiple load operation) within the evaluation period. Consequently, any full-scale experiment would be lengthy and expensive.

Inconsistency of Manufactured Bags

As described in the EPRI Fabrics Handbook (2), the bag manufacturing process involves the coordinated activities of fiberglass production, fabric weaving, fabric finishing, and bag construction. In most instances, each of these activities are performed by different companies in a horizontally integrated manufacturing process, illustrated in figure 1. For fiberglass fabric filtration products in the U. S., there are two sources of fiberglass yarns, four major fabric weavers and finishers, and several bag manufacturers. Utilities evaluating or buying new bags generally interact only with the bag manufacturers, and in turn, the bag manufacturers interact with the finishers and weavers, who in turn interact with the yarn producers. Unless a utility's procurement specification requires specific manufacturers for each step (which is very rare) the delivered product is unique in every instance and depends primarily upon the market situations that exist between the horizontally integrated manufacturers at any point in time. For example, a utility ordering replacement bags from the same bag vendor that supplied the previous set of bags (i.e., with identical specification and part number), may receive bags in which the fiberglass yarns, woven fabric, and finish were supplied by entirely different companies. The problem is further complicated because there are no universally accepted tests for confirming the consistency and quality of the delivered products (3). The impact of product inconsistency on full-scale baghouse performance is very difficult to track, and to date has been tolerated by baghouse operators. However, when evaluating the relative performance of new fabrics and bags the manufacturing history must be known in detail.

Proprietary Features of Fabrics

As discussed above, woven fiberglass fabrics used to manufacture filter bags are available from several U. S. vendors. In general, each vendor offers an equivalent product in terms of commodity properties (e.g., weave pattern, yarn count, yarn construction, materials of construction). What distinguishes one fabric from another is the finish and the manufacturing process for applying the finish - the proprietary features of a woven fiberglass fabric. This situation complicates the evaluation of different filtration fabrics because the evaluator typically cannot independently control the specification or consistency of the proprietary properties, which heretofore have been regarded as major factors in the performance of specific filter bags.

No Accepted Methodology for Evaluating Bags

As one may conclude from the discussion above, there has not been any universally acceptable method for screening or evaluating the performance of bags in fabric filters. To further complicate the situation, fabric tests such as MIT-Flex and Mullen-burst are controversial and cannot be relied upon for fabric quality control, fabric acceptance testing. or filter bag life evaluations (2,3).

EXPERIMENTAL APPROACH

The methodology described below overcomes the historical problems discussed above and has been used to effectively evaluate the relative performance of different bags. The experimental approach that was adopted assumes that performance characterizations of different bags in the same compartment will produce valid comparative information, providing that only one fabric parameter (e.g., fabric finish) is varied per compartment, and that drag and dustcake weight can be monitored for individual bags.

PERFORMANCE PARAMETERS

The performance of a particular fabric (or bag) in a baghouse is measured four ways: (1) the resistance to flow, or drag, that develops as the dustcake grows; (2) residual dustcake weight that increases with hours of operation; (3) the particulate matter collection efficiency; and (4) time to catastrophic failure of the bag. For purposes of this experiment it is assumed that since it is primarily the residual dustcake which filters particulate matter from the gas stream, the collection efficiency is satisfactory for all fabrics. Further, practical considerations for the length of the overall experiment suggest that continuing an experimental run until catastrophic failure is observed is not justified. Hence, the experimental method utilizes drag and dustcake weight as measurements of the principal performance parameters in the comparative evaluation of different bags.

Average Drag

Drag is the pressure drop across a bag normalized by the flow through the bag. Drag is typically defined as the ratio of tubesheet pressure drop in inches of H_2O to air-to-cloth ratio in acfm/ft^2, or equivalently, ft/min. The units of drag are in. H_2O min/ft. Drag can be determined instantaneously or as an average. For this experiment, average drag, or the drag midway through a filtration period (from the time a compartment comes back on-line after cleaning to the time the compartment is taken off-line for cleaning) was determined for each test fabric and was used to rank drag performance.

Dustcake Weight

Dustcake weight is the areal density of the residual dustcake, measured in lb/ft^2. Dustcake weight is determined by dividing the weight of the residual dustcake in a test bag (measured bag weight minus the weight of a clean bag) by the filtering surface area inside the bag (49.6 ft^2). In this experiment, bags were weighed at approximate 1000-h intervals.

PILOT PLANT WITH FULL-SCALE BAGS

The preferred approach for evaluating fabrics and bags is to utilize a fabric filter pilot plant in which the operating conditions can be well controlled, and which utilizes full-scale bags. The EPRI 2.5 MW fabric filter pilot plant at the

Arapahoe Test Facility is a facility ideally suited for this purpose. As summarized in Table 1, the pilot unit has four compartments, 36 bags per compartment, and uses 8 in. diameter, 23.7 ft long bags. The general layout of the pilot plant is shown in figure 2.

MEASURING GAS FLOW THROUGH INDIVIDUAL BAGS

At any time during a filtration period, pressure drop is essentially equal across each bag in the pilot plant. However, the rate of flow of flue gas through each bag (i.e., air-to-cloth ratio) is not equal but varies according to the resistance to flow of the fabric and dustcake. Conventional measurements of compartment pressure drop and air-to-cloth ratio do not provide for meaningful bag-to-bag comparisons. Hence, it is critical to monitor the gas flow rates through individual bags. For this purpose, an Individual Bag Flow Monitoring (IBFM) system, manufactured by ETS, Inc.(4), was installed in the EPRI 2.5 MW fabric filter pilot plant. Each of the 144 thimbles in the four tubesheets were equipped with IBFM flow sensors. The instrumentation package permits on-line monitoring of one-fourth (i.e., 36 bags) at a time. The system is shown schematically in Figure 3.

The IBFM system described herein is mechanically and electronically complex, generates significant volumes of data, and requires regular maintenance attention. Nonetheless, it adequately served a critical need for the experiment. Figure 4 is an example of the data produced from the IBFM system. In this example the air-to-cloth ratio was measured for each of 36 bags in a compartment and averaged over 18 filtration periods (one filtration period lasted 3 h). These data were taken 1500 hours after the installation of 2 new bags in this compartment; the other 34 bags had fully-developed dustcakes from a prior experiment and had been in service for over one year. The old bags and the new bags were identical. The old bags show remarkably uniform flow characteristics; average air-to-cloth ratios were 1.82 + 0.04 ft/min. The 2 new bags had identical air-to-cloth ratios of 2.66 + 0.06 ft/min. Of particular interest is that the new bags were taking 40% more gas flow than the old bags after 1500 hours of operation. Although, not specifically documented here, the air-to-cloth ratios for the new bags immediately after installation were on the order of 10 ft/min. These observations quantitatively illustrate the compensating operating character of baghouses for old and new bags.

TEST CONFIGURATION

There were numerous bag fabrics and operational test parameters which could be investigated. Since this experiment was planned to last for at least 5000 h of flue gas exposure, fabric design parameters had to be prioritized before the test bags could be procured.

The initial test program concentrated on woven fiberglass bags, the conventional choice for U. S. utility baghouses. Accordingly, the fiberglass design and manufacturing parameters listed in Table 2 were considered for investigation.

Based on priorities established in cooperation with the Manufacturers Advisory Group, the following four parameters were selected for initial investigation: yarn construction, texturization, finish, and weave. Each of the four compartments in the 2.5 MW pilot plant were equipped with bags that were equivalent except for variations in the parameter of interest. Also, a "reference fabric" was defined for the test. This reference fabric was chosen based on the most common bag fabric and finish installed in utility baghouses and is described in Table 3. Bags made from the reference fabric were installed in each compartment; six reference bags were installed in compartments 1 and 2, and four reference bags were installed in compartments 3 and 4. The reference test bags described in Table 3 were used as the basis of comparison for evaluating the performance of the test bags and permitted additional comparisons to be made between bags in different compartments. The general layout for the initial test configuration is shown in Figure 5.

Test Matrix for the Initial Screening of Fabric Design Parameters

Each compartment of the 2.5 MW pilot baghouse was devoted to one of the specific fabric parameters selected for evaluation. Compartment 1 was devoted to yarn construction, compartment 2 to texturization, compartment 3 to finish, and compartment 4 to weave. The complete matrix summarizing the individual test bags is presented in Table 4.

In the yarn-construction compartment, warp yarn constructions included both singles and plied filament yarns. Fill yarn constructions included texturized singles, texturized plied, and filament singles plied with texturized singles yarns. Yarns with lower twists were also included. In the texturization compartment, all fabrics were woven in a 3 x 1 twill and different levels of exposed surface texturization (EST) were evaluated. EST is varied from zero to 100% by selection of texturized or non-texturized yarns and the fabric surface orientation in bag sewing. The degree of fill yarn texturization (FYT), indicated by the fill yarn yield, was also varied from 3400 to 3700 yd/lb. In the finish compartment, different proprietary finishes applied to heat cleaned reference fabric were evaluated. The reference fabric was all woven by the same weaver, and Burlington, Stevens, Uniglass, Menardi-Criswell, and W. L. Gore each applied their respective finish to the fabric. In the weave compartment, a variety of fabric weaves with a common Teflon B finish was evaluated. Wherever possible, the yarn count and yarn construction were that of the reference fabric. In addition, a highly texturized 17 oz/yd^2 fabric and a lightweight untexturized, high yarn count fabric were tested. For comparison purposes, light weight Ryton needlefelt, woven Ryton, and light weight Tefaire needlefelt bags were also tested in the weave compartment.

The results of a 12-month evaluation of these test bags is presented in the companion paper of Felix et al. (1).

The work described in this paper was not funded by the U. S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

ACKNOWLEDGMENTS

This work was sponsored by the Electric Power Research Institute under project RP1129-8. The EPRI project managers for this work were Mr. Walter V. Piulle and Dr. Ramsay L. Chang. Their support and guidance are appreciated. The authors also wish to acknowledge the contributions of the companies and individuals who have made this investigation possible. In particular we wish to thank Mr. Theron Grubb of Grubb Filtration Testing Services, Inc. who assisted in the prioritization of test parameters and procurement of the test fabrics, the participating companies in the Manufacturers Advisory Group who worked together to assure an objective technical evaluation and who contributing their time and resources to fabrication of the test bags, Mr. Robert Carr, formerly Director of the Environmental Control System Department at EPRI and founder of EPRI's Fabric Filter Research Program for supporting the development of this test methodology, and the members of an the ad-hoc utility advisory committee to EPRI for fabrics research: Mr. Noel Wagner of the Pennsylvania Power and Light Company, Ms. Cheryl Cortright of the Salt River Project, Mr. Robert Gehri of Southern Company Services, and Mr. Tim Conkin of the Los Angeles Department of Water and Power.

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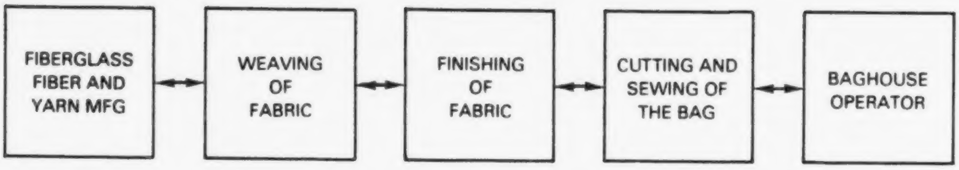


Figure 1. Horizontally integrated manufacturing process for production of fiberglass fabric filter bags.

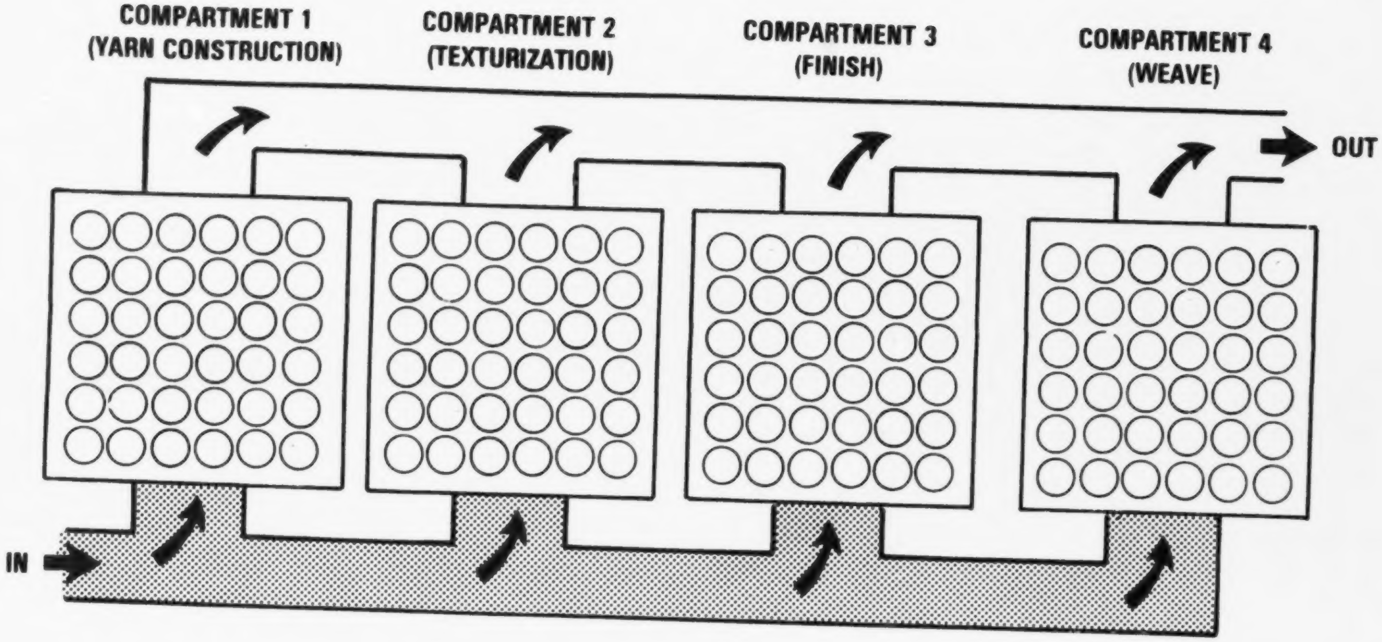


Figure 2. EPRI 2.5 MW pilot fabric filter.

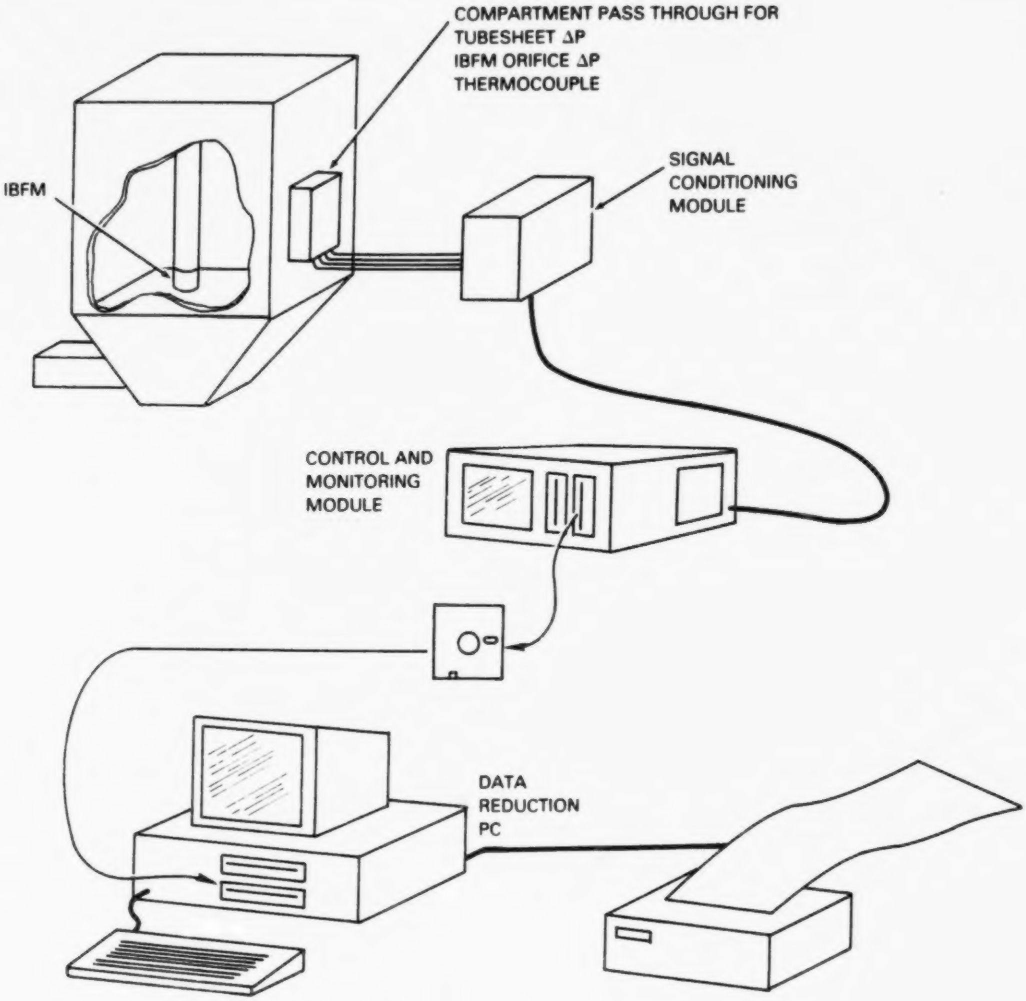
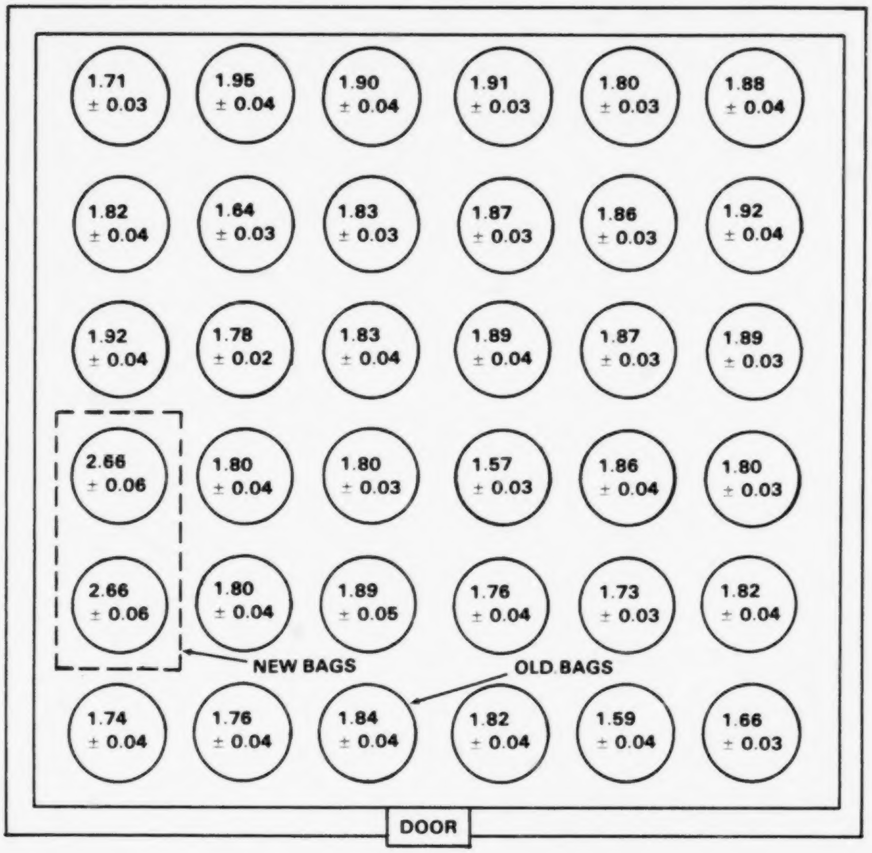


Figure 3. IBFM system for monitoring drag and flow through individual filter bags.



74 DAYS AFTER
NEW BAGS INSTALLED

Figure 4. Average air-to-cloth ratios (in ft/min) measured with the IBFM system in one compartment of the EPRI 2.5 MW pilot baghouse.

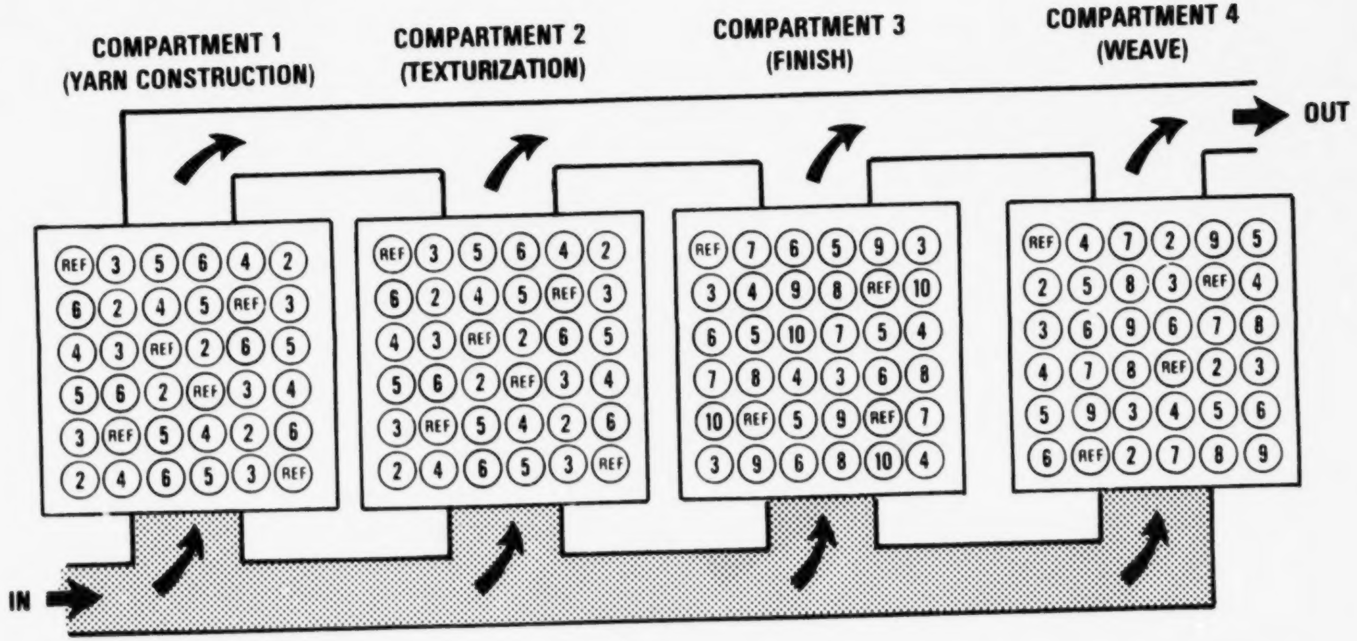


Figure 5. Initial test configuration for evaluation of different bags in the EPRI 2.5 MW pilot baghouse.

TABLE 1. DESIGN AND OPERATING PARAMETERS FOR THE EPRI 2.5 MW PILOT FABRIC FILTER AT THE ARAPAHOE TEST FACILITY

Number of Compartments	4
Number of Bags per Compartment	36
Bag Diameter (inches)	8
Bag Length (feet)	23.7
Bag Area (ft^2)	49.6
Flue Gas Flow Rate (acfm)	10,000
Inlet Temperature (°F)	270
Gross Filtering A/C (acfm/ft^2)	1.5
Reverse Gas Flow Rate (acfm)	3,500
Reverse Gas A/C (acfm/ft^2)	2.0
Bag Tension (lb)	60
Filtering Time between Cleans (min)	180
Reverse Gas Clean Timing Parameters	
Null (seconds)	15
Reverse Gas (seconds)	30
Null (seconds)	30

TABLE 2. WOVEN FIBERGLASS FABRIC DESIGN AND MANUFACTURING PARAMETERS

Yarn Count	The number of warp and fill yarns per square inch of fabric.
Yarn Construction	How the fiberglass strands are gathered, twisted, and plied together.
Strand Weight	The number of fiberglass filaments per yarn.
Weave Pattern	3 x 1 Twill, 2 x 2 Twill, Crowfoot, etc.
Finish	Proprietary (e.g., Teflon B, Acid Resistant, etc.).
Coronizing	Heat treatment after weaving.
Yarn Type	Texturized or continuous filament.
Filament Diameter	Diameter of a single fiberglass strand.
Material Composition	"E-glass", "S-glass", etc.

TABLE 3. SPECIFICATIONS OF THE REFERENCE TEST BAG FABRIC

Warp Yarn	150 1/2 4.0Z/3.8S
Fill Yarn	150 1/4 TEXT 4.0Z/3.8S
Weave	3 x 1 Right Hand Twill
Yarn Count	54 x 30 /in., warp x fill
Material	ECDE fiberglass
Finish	10% Teflon B
Fabric Orientation	Warp out
Weight	10 oz/yd^2
Fill Yield	3400 yd/lb

TABLE 3. SPECIFICATIONS OF THE REFERENCE TEST BAG FABRIC

Warp Yarn	150 1/2 4.0Z/3.8S
Fill Yarn	150 1/4 TEXT 4.0Z/3.8S
Weave	3 x 1 Right Hand Twill
Yarn Count	54 x 30 /in., warp x fill
Material	ECDE fiberglass
Finish	10% Teflon B
Fabric Orientation	Warp out
Weight	10 oz/yd^2
Fill Yield	3400 yd/lb

TABLE 4. DESCRIPTIONS OF FABRICS TESTED

Test Parameter	Compartment	Fabric ID	Warp Yarn Designation	Fill Yarn Designation	Weave	Yarn Count	Yarn Material	Finish	Warp In/Out	Source of Tex. Yarn	Approx. Fab. Wt. (oz/yd^2)	Fill Approx. Yield (yd/lb)
Yarn Construction	1	1-1	(Ref.) 150 1/2-4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW[a]	54x30	ECDE	MS-10% Teflon B	Out	Weaver	10	3400
Yarn Construction	1	1-2	75 1/0 0.7Z	37 1/0 TEX 1.0Z	1x3 RHTW	54x30	ECDE	MS-10% Teflon B	Out	Glass Mfr.	10	3400
Yarn Construction	1	1-3	75 1/0 0.7Z	150 1/0 1.0Z + 50 1/0 TEX 1.0Z/2.5S	1x3 RHTW	54x30	ECDE	MS-10% Teflon B	Out	Glass Mfr.	10	3400
Yarn Construction	1	1-4	75 1/0 0.7Z	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	MS-10% Teflon B	Out	Weaver	10	3400
Yarn Construction	1	1-5	150 1/2 4.0Z/3.8S	75 1/2 TEX 1.0Z/2.5S	1x3 RHTW	54x30	ECDE	MS-10% Teflon B	Out	Glass Mfr.	10	3400
Yarn Construction	1	1-6	150 1/2 3.0Z/2.8S	150 1/4 TEX 3.0Z/2.8S	4 HS[b]	54x30	ECDE	MS-10% Teflon B	Out	Weaver	10	3400
Texturization	2	2-1	(Ref.) 150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	MS-10% Teflon B	Out	Weaver	10	3400
Texturization	2	2-2	150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	MS-10% Teflon B	Out	Weaver	10	3600
Texturization	2	2-3	150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	MS-10% Teflon B	Out	Weaver	10	3200
Texturization	2	2-4	150 1/2 4.0Z/3.8S	150 1/4 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	MS-10% Teflon B	Out	N/A	10	3700
Texturization	2	2-5	150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	MS-10% Teflon B	In	Weaver	10	3400
Texturization	2	2-6	150 1/2 4.0Z/3.8S	150 1/2 4.0Z/3.8S	1x3 RHTW	54x56	ECDE	MS-10% Teflon B	Out	Napped	8.8	7500
Finish	3	3-2	(Ref.) 150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	MS-10% Teflon B	Out	Weaver	10	3400
Finish	3	3-3	150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	MS-Gore-Tex[c]	Out	Weaver	10	3400
Finish	3	3-4	150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	UM-7% Teflon B	Out	Weaver	10	3400
Finish	3	3-5	150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	MS-XX	Out	Weaver	10	3400
Finish	3	3-6	150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	UM-YY	Out	Weaver	10	3400
Finish	3	3-7	150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	JPS-ZZ	Out	Weaver	10	3400
Finish	3	3-8	37 1/0 1.0Z	Fiberglass-PTFE blend	1x3 RHTW	47x24	ECDE	JPS-AA	Out	Weaver	14	-
Finish	3	3-9	150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	BGF-BB	Out	Weaver	10	3400
Finish	3	3-10	150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	BGF-CC	Out	Weaver	10	3400
Weave	4	4-1	(Ref.) 150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	MS-10% Teflon B	Out	Weaver	10	3400
Weave	4	4-2	150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	4 HS[b]	54x30	ECDE	MS-10% Teflon B	Out	Weaver	10	3400
Weave	4	4-3	150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	2x2 BKNTW[d]	54x30	ECDE	MS-10% Teflon B	N/A[e]	Weaver	10	3400
Weave	4	4-4	37 1/0 1.0Z	(1)75 1/0 0.7Z + (2)75 1/0 TEX 1.0Z/2.5S	1x3 RHTW	44x24	ECDE	MS-10% Teflon B	Out	Glass Mfr.	14	2300
Weave	4	4-5	37 1/0 1.0Z(ALT) 37 1/0 TEX 1.0Z	75 1/3 TEX 4.0Z/3.8S	Special[f]	48x30	ECDE	MS-10% Teflon B	Fil. Out[g] Tex. In	Weaver	17	2300
Weave	4	4-6	Restex scrim	Ryton needlefelt	Felt	N/A	Ryton	N/A	N/A	N/A	13	N/A
Weave	4	4-7	18.5/1 C.C.[h]	16/1 C.C.[h]	3x2 TW	64x60	Ryton	Heat-set	Out	N/A	6	13,400
Weave	4	4-8	Dupont Tetaire[i]	-	Felt	N/A	Tetaire	N/A	N/A	N/A	13	N/A
Weave	4	4-9	450 1/0 1.0Z	100 1/0 1.0Z	4 HS[b]	100x60	ECD(warp) ECDE(fill)	BGF-6% Teflon B	Out	N/A	5	10,000

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[a] Right-hand twill.

[b] 4 Harness satin.

[c] MS-10% Teflon B with Gore-Tex PTFE laminate on filling face.

[d] Broken twill.

[e] Symmetrical weave.

[f] Double crowfoot.

[g] Filament warp out/texturized warp in.

[h] C.C. = Cotton count (spun yarns) = number of 840-yd lengths per pound of yarn.

[i] Tetaire is a blend of Teflon PTFE fibers and ECB glass fibers needled onto a woven Teflon scrim.

PARAMETRIC TESTING OF FILTRATION FABRICS FOR LOW-SULFUR-WESTERN COAL FLY ASH

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ABSTRACT

This paper reports the results of a filtration fabric evaluation program conducted at the EPRI Arapahoe Test Facility in Denver, CO. In this test, 22 ft long, 8 in. diameter filter bags made from 27 different filtration fabrics were tested for over one year in a 2.5-MW(e), four-compartment, pilot-scale baghouse. The baghouse was cleaned with reverse gas. Filtering air-to-cloth ratios ranged from 1.5 to 2.0 acfm/ft^2. Special instrumentation allowed the flow through any bag to be monitored throughout a filtering cycle.

One fabric related parameter was tested in each compartment: yarn construction (compartment 1), yarn texturization (compartment 2), finish (compartment 3), and weave (compartment 4). Bags made from a reference fabric were also installed in each compartment.

Fabric finish and weave did not appear to affect performance (drag and dustcake weight). However, a strong correlation between fabric surface texturization and dustcake weight was observed.

INTRODUCTION

Electric utilities have made significant progress in recent years designing and operating bag filters for particulate collection. As a result of these advances, baghouses have become an accepted and frequently preferred particulate control approach. The challenge now is to continue to improve baghouse performance and reduce the cost of electricity.

Although utility baghouse performance is known to be coal specific, a major determinant of baghouse performance is the fabric from which bags are made (1,2). The effect of fabric variables such as yarn construction, weave, finish, and texturization on dustcake weight, drag, emissions, and ultimately bag life is not well understood. Consequently, there is no defined method for determining the best fabric for a given application.

This paper presents and discusses results of an extensive fabrics evaluation program at the EPRI Arapahoe Test Facility (ATF) in Denver, CO (3). The goal of this evaluation is to determine which parameters of fabric construction and finish most influence filter bag performance for low-sulfur, western subbituminous coal fly ash. Because a large number of fabric types were tested in this evaluation, multiple types of fabric were installed on each compartment of the test baghouse. Therefore, emissions associated with a particular type of test fabric could not be measured, and average drag (drag averaged between compartment cleanings) dustcake weight (areal density of the residual dustcake) were used to rank bag performance. Emissions information was not significant because previous testing at the ATF has shown that emissions were always more than an order of magnitude less than limits permitted by the New Source Performance Standards (NSPS) even for fabrics with the poorest collection efficiencies (3). Also, because of the relatively short length of this test no study of bag life was possible.

The fabrics and bags for this study were provided at no cost by members of the EPRI Manufacturers Advisory Group (MAG). This group includes makers and weavers of fiberglass and non-fiberglass fibers, fiberglass finishers and finish suppliers, bag fabricators, and utilities. The following are the companies that are represented in this group:

BHA Group, Inc.
Burlington Glass Fabrics Company
E. I. du Pont de Nemours & Company, Inc.
J. P. Stevens & Company, Inc.
Menardi-Criswell
Midwesco, Filter Media Resources Division
Owens-Corning Fiberglas Corporation
PPG Industries, Inc.
Tex Tech Industries, Inc.
Uniglass Industries (Division of United Merchants and Manufacturers, Inc.)
W. L. Gore & Associates, Inc.

The following EPRI utility members are also represented in this group:

City of Los Angeles Department of Water and Power
Pennsylvania Power and Light Company
Salt River Project
Southern Company Services, Inc.

The ATF is associated with the Arapahoe station operated by the Public Service Company of Colorado (PSCCO). The ATF operates on a flue gas sidestream from the Arapahoe unit 4 boiler that is fueled with a low-sulfur, western subbituminous coal from Wyoming. Typical coal and ash analyses are shown in table 1.

The fabrics evaluation discussed in this paper was carried out at the ATF 2.5-MW(e) pilot-scale baghouse. This four-compartment baghouse has 36 bags per compartment. The bags are 22 ft long and have a diameter of 8 in. (a surface area of 49.6 ft^2). Each compartment can employ reverse gas, reverse gas with sonic assist, or shake/deflate cleaning. Although both reverse-gas cleaning and sonically assisted reverse-gas cleaning was tested, for the majority of this evaluation reverse-gas cleaning was used. Therefore, this paper will focus on results obtained with this cleaning method. Filtering air-to-cloth ratios ranged from 1.5 to 2.0 $acfm/ft^2$. Table 2 lists pertinent design and operating parameters for this baghouse.

Each bag in the baghouse was equipped with an Individual Bag Flow Monitor (IBFM) manufactured by ETS, Inc. (3,4). With this instrument, the flow through a bag can be monitored while it is in service. Drag information obtained with the IBFM's and periodic measurement of residual dustcake weights were the principal data available for each of the fabrics tested in this evaluation. In addition to the IBFM's, each compartment in the pilot-scale baghouse was fitted with a large orifice between the compartment and the outlet poppet valve. These orifices were calibrated during air-load tests and were used to check the IBFM flow measurements.

Throughout this test, the unit 4 ATF host boiler was used for voltage stabilization in the metropolitan Denver area. Thus, boiler loads were rarely steady and high at the same time. Virtually all steady load operation was at relatively low boiler loads. Therefore, in order to study how average drag evolved for each of the test fabrics over the duration of this test, only low load data could be included.

DESCRIPTION OF THE TEST FABRICS

One compartment of the baghouse was devoted to each major parameter of filter fabric construction or finish (yarn construction, texturization, finish, and weave). In addition, bags made from a "reference" fabric were installed in each compartment to facilitate the comparison of bag performance between compartments. The reference fabric is a "plied-plied" (plied yarns in both the warp and fill directions), 1 x 3 right hand twill weave, 10 oz/yd^2 weight, fiberglass fabric that was heat cleaned and coated with a 10% Du Pont Teflon B finish. This reference fabric was selected because it is widely used in utility baghouses. Table 3 provides a complete description of the fabrics tested in this evaluation. The particular reference fabric used for this evaluation was chosen because it is widely used in utility baghouses.

In compartment one, the effect of yarn construction on drag and dustcake weight was evaluated. Warp yarn constructions included both singles and plied filament yarns, and fill yarn constructions included texturized singles, texturized plied, and filament singles plied with texturized singles yarns. Yarns with lower levels of twist (Fabric 1-6) were also included.

In compartment 2, the effect of fabric surface texturization was evaluated. All fabrics were woven in a 1 x 3 right hand twill and different levels of exposed surface texturization (EST) were obtained by exposing different sides of the test fabric to the dirty flue gas, by varying the filling yarn texturization (FYT) and by napping (mechanically roughening the filling surface of a nontexturized fabric).

A typical 1 x 3 twill fabric has approximately 75% of its surface covered with texturized yarns on the side of the fabric on which the texturized fill yarns predominate (the filling side). For maximum fabric strength, only fill yarns are texturized. The filling side usually faces the dirty gas stream, and when bags are made in this manner they are described as having a "warp-out" construction or 75% EST. If the fabric is reversed so that the warp side of the fabric faces the dirty gas stream, the bag is said to have a "warp-in" construction or 25% EST (fabric 2-5). If the filling yarns are not texturized (filament yarns), the fabric has 0% EST (fabric 2-4).

In compartment 3, fabric finish was the primary variable. With the exception of fabrics 3-3 and 3-8, all of the test fabrics in compartment 3 were heat cleaned reference fabric to which a variety of finishes had been applied, including: 10% and 7% Teflon B, and acid resistant and experimental finishes. Fabric 3-3 was the reference fabric with a 10% Teflon B finish to which a GORE-TEX membrane had been laminated. Fabric 3-8 was an experimental fabric with a combination fiberglass and polytetrafluoroethylene (PTFE) fill yarn.

In compartment 4 a variety of fabric weaves was evaluated along with two nonfiberglass fabrics. All of the fiberglass fabrics tested in compartment 4 were finished with a 10% coating by weight of Teflon B. The same yarn construction and yarn count were used for three fabrics: a 1 x 3 right hand twill (the reference fabric, 4-1), a four-harness satin weave (fabric 4-2), and a 2 x 2 broken twill weave (fabric 4-3). The other three fiberglass fabrics in this compartment were a 14 oz/yd^2 fabric that is typical of the heavier fabrics that are sometimes used in utility baghouses (fabric 4-4), a 17 oz/yd^2 "double crowfoot" weave fabric with filament and texturized warp yarns and a texturized fill yarn (fabric 4-5), and a very high count, four-harness satin weave fabric made with nontexturized fill yarns (fabric 4-9). This "fine filament" fabric had a very smooth surface. Because of the different weaves, EST was also a variable in compartment 4.

The nonfiberglass fabrics included a light weight Ryton needlefelt fabric with a RASTEX scrim (fabric 4-6), a woven Ryton fabric made from staple Ryton yarns in a 3 x 2 twill (fabric 4-7), and a lightweight Tefaire needlefelt fabric (fabric 4-8). Ryton is Phillips Fiber's polyphenlyene sulfide material. RASTEX is the W. L. Gore Company's proprietary yarn made from slit GORE-TEX membrane. Tefaire is Du Pont's proprietary felted fabric made from a blend of fiberglass and Teflon fibers. The two light weight felts were much more pliable and were not as dense as the versions of these fabrics normally used in pulse-jet cleaned baghouses.

EXPERIMENTAL RESULTS

The fabrics evaluation program with reverse-gas cleaning lasted for 6230 h of flue gas exposure. Throughout this test, every attempt was made to operate the pilot-scale baghouse as though it were a full-scale unit. The one exception to this operating plan was when the Arapahoe unit 4 host boiler was occasionally fired with natural gas during the winter of 1986-87. The pilot-scale baghouse was brought off line during these periods while the unit 4 baghouse was not brought off line.

Figure 1 shows how the reference fabric in each of the four compartments behaved. In this graph, average drag (drag averaged between compartment cleanings) is plotted against time in service, including a short period of sonically augmented reverse-gas cleaning at the end of testing. Although results obtained for sonically augmented reverse-gas cleaning are not discussed in this paper, these data are included in this figure to show the reduction in average drag that is possible with more energetic cleaning. Each data point in figure 1 represents an average of approximately three calendar weeks of data. This graph indicates that bags made from the reference fabric behaved similarly throughout the test, regardless of the compartment in which they were installed. Figure 1 also shows that at least four thousand hours of dervice were required before the filtration performance of these reverse-gas cleaned bags had stabilized.

The relative average drag and dustcake areal density of the fabrics tested in the different compartments are shown in figures 2 through 5. Relative average drag was calculated by normalizing the average drag behavior of the bags of each type of fabric in each compartment to that of the reference fabric bags in that compartment. Dustcake weights were measured on a periodic basis by weighing all of the bags in every compartment and calculating average bag weights and dustcake areal densities for each type of bag. Results for each compartment are summarized below:

COMPARTMENT 1 - YARN CONSTRUCTION

- Relative to the reference fabric, fabrics with singles yarns in the warp and fill (fabrics 1-2 and 1-3) had about 12% higher average drags and from 14% to 18% higher dustcake weights.

- Fabric 1-6, that was woven with a lower than average twist in the warp and fill yarns had 3% lower average drag and 9% lower dustcake weight than the reference fabric.

- Other variations in yarn construction did not produce significant changes in average drag or dustcake weight.

These results suggest that fabrics woven from plied warp and fill yarns tend to perform slightly better than fabrics woven from singles warp and fill yarns. Also, fabrics woven with filling yarns that have slightly less twist than is usual also perform slightly better than the reference fabric. As will be seen from the behavior of the test fabrics in the compartment 2, the improved performance of the fabric 1-6 compared to the reference fabric probably was due to its FYT being slightly lower than that of the reference fabric.

COMPARTMENT 2 - TEXTURIZATION

- EST significantly affected dustcake weight and drag. The all filament 0% EST bag (fabric 2-4) had 21% lower average drag and 56% less dustcake weight than the reference fabric bags that had 75% EST. The 25% EST bag (fabric 2-5) had 14% lower average drag and 25% less dustcake weight than the reference fabric.

- Changes in FYT (fabrics 2-2 and 2-3) produced minor changes in performance, compared to performance changes caused by altering FYT.

- The napped fabric (fabric 2-6) behaved like the 75% EST reference fabric.

COMPARTMENT 3 - FINISH

- Fabric finish has no effect on average drag or dustcake weight.

- The experimental fabric with a blended fiberglass and PTFE filling yarn (fabric 3-8) had an average drag equal to that of the reference fabric but with a 31% lighter dustcake weight. The superior performance of fabric 3-8 may be due to a filling yarn that is smoother than that the texturized filling yarns of the other fabrics in compartment 3.

COMPARTMENT 4 - WEAVE

- Weave has no effect on filter performance for fiberglass fabrics with levels of EST greater than 50%. The 1 x 3 twill weave fabric (fabric 4-1), the four-harness satin fabric (fabric 4-2), and the heavier, 14 oz/yd^2, 1 x 3 twill fabric (fabric 4-4) all had nearly equal average drag and dustcake weights. These fabrics had the same EST, with their filling faces (75% EST) exposed to the dust. The 2 x 2 broken twill fabric, with 50% EST (fabric 4-3) and the 100% EST heavy double crowfoot weave fabric (fabric 4-5) also behaved much like the reference fabric. These results indicate that higher levels of EST (>50%) do not influence average drag and dustcake weight as strongly as lower levels of EST (<50%).

- Lightweight felted fabrics are unsuitable for reverse-gas cleaned baghouses. All of the felted fabrics had abnormally high average drag and dustcake weights. The light weight Tefaire bags (fabric 4-8) also stretched more than one ft.

- Woven Ryton fabric made from staple Ryton yarns (fabric 4-7) is unsuitable for reverse-gas cleaned baghouses due to unacceptably high average drag and dustcake weight. The exterior of these bags had an almost flannel-like feel and appearance. If this fabric were singed or woven from continuous filament yarns, it would have a much smoother surface with a greater tendency to shed dust.

- The fine filament/high weave count experimental fabric (fabric 4-9) produced dramatic reductions in average drag and dustcake weight (11% and 67%, respectively, below the reference bag).

CONCLUSIONS

The results of this pilot-scale evaluation indicate that fabric construction parameters can be modified to improve filtration performance. While results for this study are limited to low-sulfur, Western coals, some of the concepts should be applicable to other coal types as well. A parallel program to test current findings on high sulfur coal is being conducted at EPRI's fabric filter pilot plant at the Scholz station operated by Gulf Power (Southern Company Services, Inc.) in Sneads, FL.

The surface texturization of the filtering surface of a filter fabric (EST) has more effect on average drag and dustcake weight than any other fabric construction or finish related parameter. For reverse-gas cleaned bags, lowering the degree of EST on the filtering side of the bag significantly lowers average drag and dustcake weight, especially for EST's of less than 50%.

Varying the degree of texturization of the yarn itself (FYT) produces only minor variations in fabric performance.

Fabric finish and weave does not affect average drag or dustcake weight for reverse-gas cleaned baghouses filtering low-sulfur, Western coal fly ash.

High yarn count fine filament fabrics and fabrics with composite yarns formed from a mixture of fiberglass and PTFE yarns are promising candidates for reducing average drag and dustcake weight.

Light weight felted fabrics are unsuitable for reverse-gas cleaned baghouses unless the highly texturized filtering surfaces of these fabrics can be modified for better dustcake release with lower levels of cleaning energy.

Emissions could not be measured as part of this study. While the mass emissions from any of these fabrics are expected to be below NSPS with the low-sulfur coal fly ash from the Arapahoe station, other types of coal can produce noncohesive fly ash that could "bleed through" conventional filter fabrics (1,2). For these more difficult filtration applications, a suitable balance must be made between average drag, residual dustcake weight, and collection efficiency.

The effect of fabric construction on bag life is unknown and will need to be evaluated with long-term testing. There are no results from this evaluation that indicate any of the test fabrics experienced significant deterioration.

The work described in this paper was not funded by the U. S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

ACKNOWLEDGMENTS

This work was sponsored by the Electric Power Research Institute under project RP1129-8. The EPRI project managers for this work were Mr. Walter V. Piulle and Dr. Ramsay L. Chang. Their support and guidance are appreciated. We would also like to thank Mr. Theron Grubb of Grubb Filtration Testing Services, Inc. for his valuable help in coordinating the selection and manufacturing of the fabrics tested in this experiment, and the members of the MAG for their considerable contributions of time and materials without which these results could not have been obtained.

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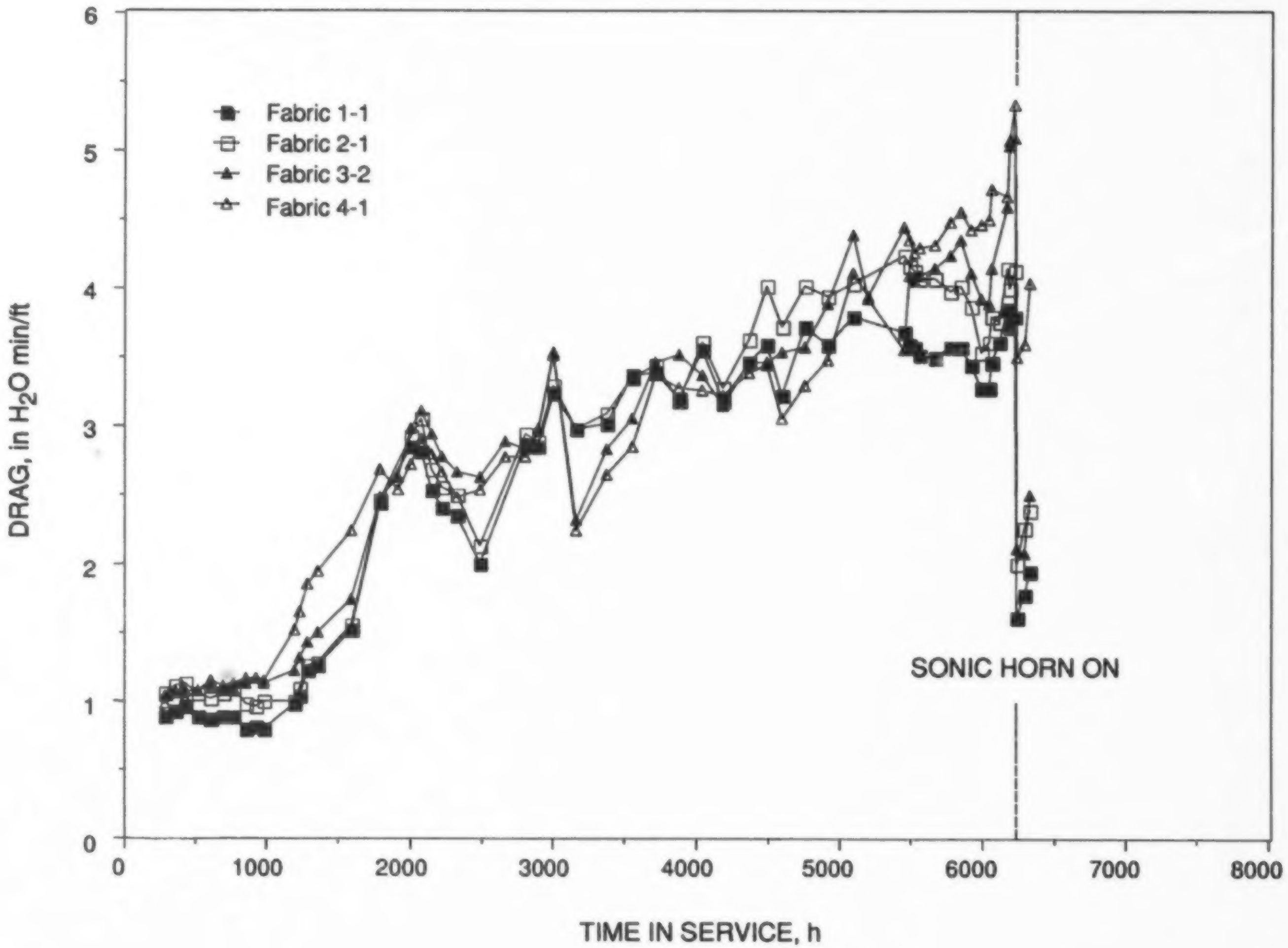

Figure 1. Average drag for the reference fabrics installed in each compartment for the duration of testing.

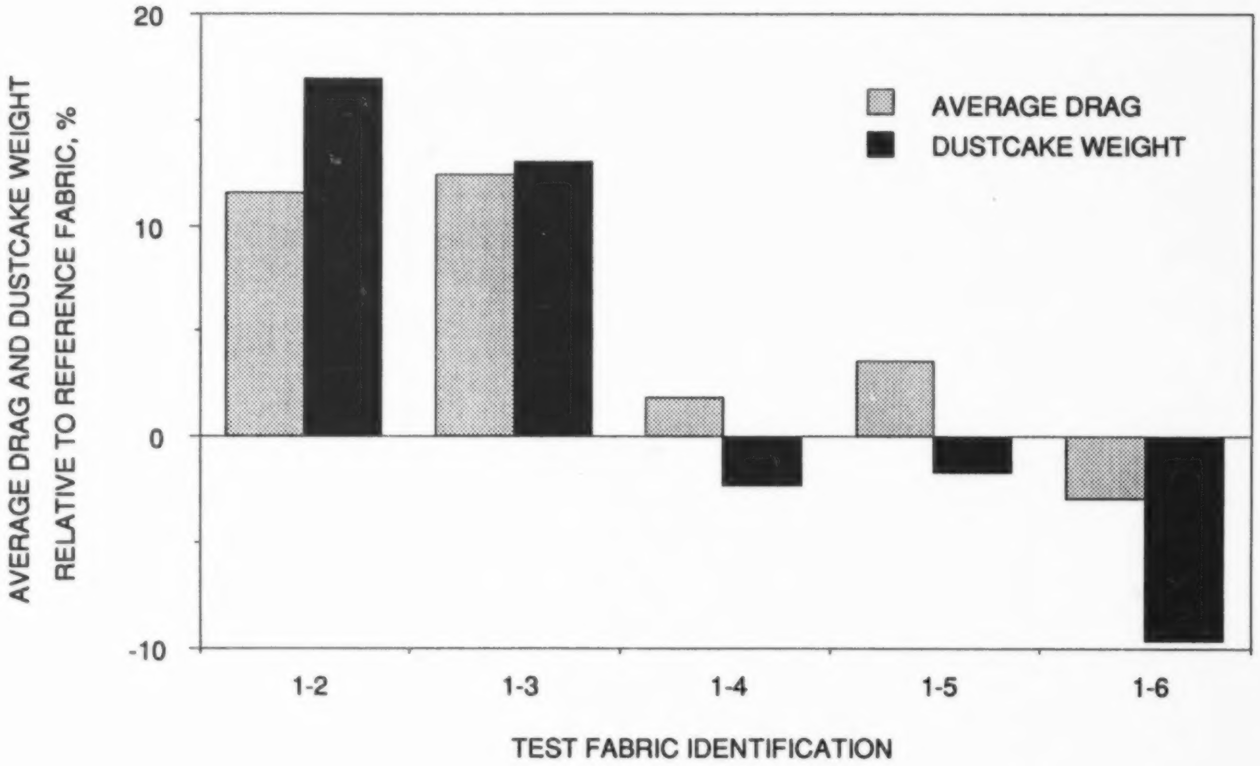

Figure 2. Average drag and dustcake weight for the test fabrics in compartment 1 relative to that of the reference fabric bags in compartment 1. Shown are averages over the time period from 5000 to 6000 h of service.

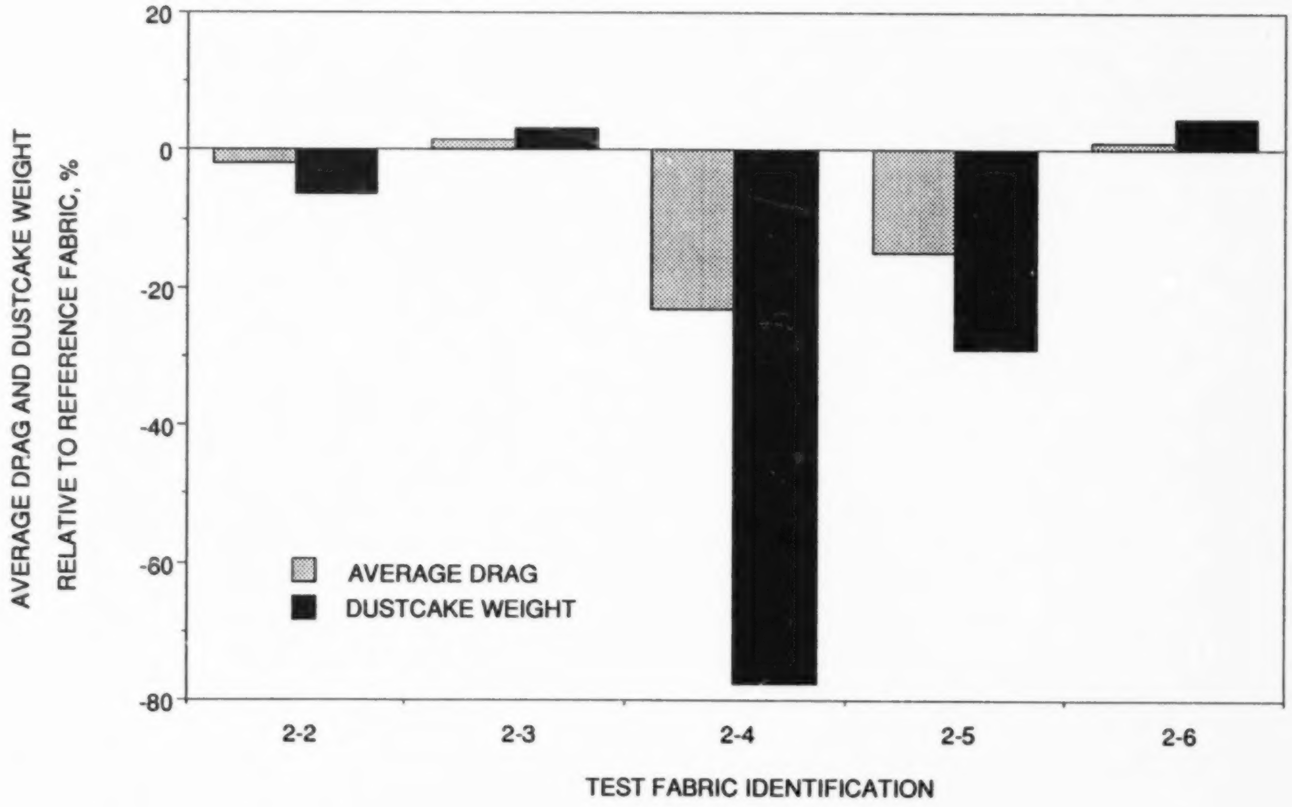

Figure 3. Average drag and dustcake weight for the test fabrics in compartment 2 relative to that of the reference fabric bags in compartment 2. Shown are averages over the time period from 5000 to 6000 h of service.

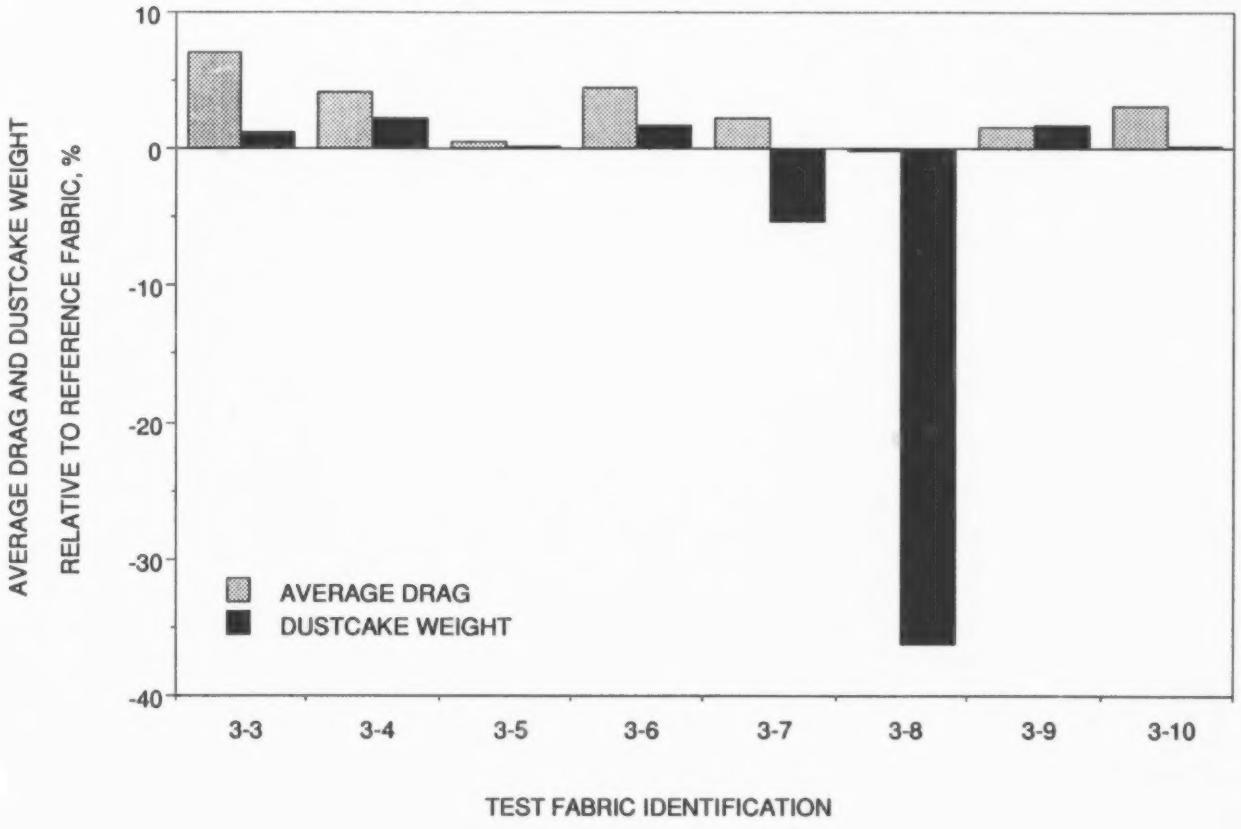

Figure 4. Average drag and dustcake weight for the test fabrics in compartment 3 relative to that of the reference fabric bags in compartment 3. Shown are averages over the time period from 5000 to 6000 h of service.

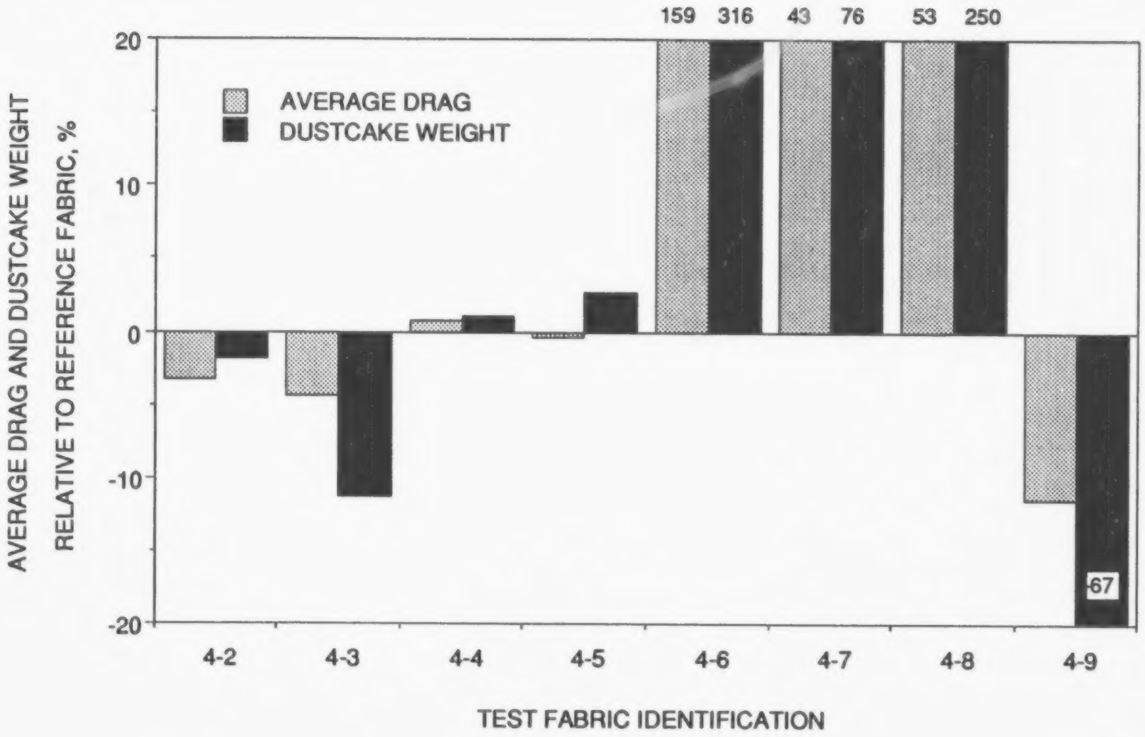

Figure 5. Average drag and dustcake weight for the test fabrics in compartment 4 relative to that of the reference fabric bags in compartment 4. Shown are averages over the time period from 5000 to 6000 h of service. The results for fabrics 4-6 through 4-9 cannot be shown with the scale chosen. Thus, the actual maximum and minimum values for those fabric types are indicated numerically above or at the bottom of the figure.

TABLE 1. COAL AND DUSTCAKE ASH ANALYSES FOR TYPICAL COAL BURNED AT THE ARAPAHOE UNIT 4 BOILER AND FLY ASH FILTERED AT THE 2.5 MW(E) PILOT-SCALE BAGHOUSE

	Coal chemistry, (% weight)	Dustcake ash chemistry, (% weight)
Na_2O	1.23	1.40
K_2O	1.12	1.30
MgO	1.34	2.00
CaO	4.75	6.20
Fe_2O_3	7.37	5.70
Al_2O_3	24.28	26.30
SiO_2	51.51	52.00
TiO_2	0.90	0.92
P_2O_5	1.09	1.60
SO_3	4.82	0.71
Sulfur	0.60	
Ash	6.76	
LiO_2		0.02
LOI		3.00
Soluble $SO_4^=$		1.80
Equilibrium Ph*		7.40

*Dimensionless.

TABLE 2. DESIGN AND OPERATING PARAMETERS FPR THE EPRI 2.5 MW(E) PILOT FABRIC FILTER AT THE ARAPAHOE TEST FACILITY

Number of Compartments	4
Number of Bags per Compartment	36
Bag Diameter (in.)	8
Bag Length (ft)	23.7
Bag Area (ft^2)	49.6
Flue Gas Flow Rate (acfm)	10,700
Inlet Temperature (°F)	270
Gross Filtering A/C Ratio (acfm/ft^2)	1.5
Reverse Gas Flow Rate (acfm)	3,500
Reverse Gas A/C Ratio (acfm/ft^2)	2.0
Bag Tension (lb)	60
Filtering Time Between Cleanings (m)	180
Reverse Gas Clean Timing Parameters: (Compartments are Batch Cleaned in Sequence)	
Null (s)	15
Reverse Gas (s)	30
Null (s)	30

Table 3. DESCRIPTION OF TEST FABRICS

Test Parameter	Compartment	Fabric ID	Warp Yarn Designation	Fill Yarn Designation	Weave	Yarn Count	Yarn Material	Finish	Warp In/Out	Source of Tex. Yarn	Approx. Fab. Wt. (oz/yd^2)	Fill Approx. Yield (yd/lb)
Yarn Construction	1	1-1	(Ref.) 150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW[a]	54x30	ECDE	MS-10% Teflon B	Out	Weaver	10	3400
Yarn Construction	1	1-2	75 1/0 0.7Z	37 1/0 TEX 1.0Z	1x3 RHTW	54x30	ECDE	MS-10% Teflon B	Out	Glass Mfr.	10	3400
Yarn Construction	1	1-3	75 1/0 0.7Z	150 1/0 1.0Z + 50 1/0 TEX 1.0Z/2.5S	1x3 RHTW	54x30	ECDE	MS-10% Teflon B	Out	Glass Mfr.	10	3400
Yarn Construction	1	1-4	75 1/0 0.7Z	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	MS-10% Teflon B	Out	Weaver	10	3400
Yarn Construction	1	1-5	150 1/2 4.0Z/3.8S	75 1/2 TEX 1.0Z/2.5S	1x3 RHTW	54x30	ECDE	MS-10% Teflon B	Out	Glass Mfr.	10	3400
Yarn Construction	1	1-6	150 1/2 3.0Z/2.8S	150 1/4 TEX 3.0Z/2.8S	4 HS[b]	54x30	ECDE	MS-10% Teflon B	Out	Weaver	10	3400
Texturization	2	2-1	(Ref.) 150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	MS-10% Teflon B	Out	Weaver	10	3400
Texturization	2	2-2	150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	MS-10% Teflon B	Out	Weaver	10	3600
Texturization	2	2-3	150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	MS-10% Teflon B	Out	Weaver	10	3200
Texturization	2	2-4	150 1/2 4.0Z/3.8S	150 1/4 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	MS-10% Teflon B	Out	N/A	10	3700
Texturization	2	2-5	150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	MS-10% Teflon B	In	Weaver	10	3400
Texturization	2	2-6	150 1/2 4.0Z/3.8S	150 1/2 4.0Z/3.8S	1x3 RHTW	54x56	ECDE	MS-10% Teflon B	Out	Napped	8.8	7500
Finish	3	3-2	(Ref.) 150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	MS-10% Teflon B	Out	Weaver	10	3400
Finish	3	3-3	150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	MS-Gore-Tex[c]	Out	Weaver	10	3400
Finish	3	3-4	150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	UM-7% Teflon B	Out	Weaver	10	3400
Finish	3	3-5	150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	MS-XX	Out	Weaver	10	3400
Finish	3	3-6	150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	UM-YY	Out	Weaver	10	3400
Finish	3	3-7	150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	JPS-ZZ	Out	Weaver	10	3400
Finish	3	3-8	37 1/0 1.0Z	Fiberglass-PTFE blend	1x3 RHTW	47x24	ECDE	JPS-AA	Out	Weaver	14	-
Finish	3	3-9	150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	BGF-BB	Out	Weaver	10	3400
Finish	3	3-10	150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	BGF-CC	Out	Weaver	10	3400
Weave	4	4-1	(Ref.) 150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	1x3 RHTW	54x30	ECDE	MS-10% Teflon B	Out	Weaver	10	3400
Weave	4	4-2	150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	4 HS[b]	54x30	ECDE	MS-10% Teflon B	Out	Weaver	10	3400
Weave	4	4-3	150 1/2 4.0Z/3.8S	150 1/4 TEX 4.0Z/3.8S	2x2 BKNTW[d]	54x30	ECDE	MS-10% Teflon B	N/A[e]	Weaver	10	3400
Weave	4	4-4	37 1/0 1.0Z	(1)75 1/0 0.7Z + (2)75 1/0 TEX 1.0Z/2.5S	1x3 RHTW	44x24	ECDE	MS-10% Teflon B	Out	Glass Mfr.	14	2300
Weave	4	4-5	37 1/0 1.0Z(ALT) 37 1/0 TEX 1.0Z	75 1/3 TEX 4.0Z/3.8S	Special[f]	48x30	ECDE	MS-10% Teflon B	Fil. Out[g] Tex. In	Weaver	17	2300
Weave	4	4-6	Rastex scrim	Ryton needlefelt	Felt	N/A	Ryton	N/A	N/A	N/A	13	N/A
Weave	4	4-7	18.5/1 C.C.[h]	16/1 C.C.[h]	3x2 TW	64x60	Ryton	Heat-set	Out	N/A	6	13,400
Weave	4	4-8	Dupont Tefaire[i]	-	Felt	N/A	Tefaire	N/A	N/A	N/A	13	N/A
Weave	4	4-9	450 1/0 1.0Z	100 1/0 1.0Z	4 HS[b]	100x60	ECD(warp) ECDE(fill)	BGF-6% Teflon B	Out	N/A	5	10,000

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[a] Right-hand twill.

[b] 4 Harness satin.

[c] MS-10% Teflon B with Gore-Tex PTFE laminate on filling face.

[d] Broken twill.

[e] Symmetrical weave.

[f] Double crowfoot.

[g] Filament warp out/texturized warp in.

[h] C.C. = Cotton count (spun yarns) = number of 840-yd lengths per pound of yarn.

[i] Tefaire is a blend of Teflon PTFE fibers and ECB glass fibers needled onto a woven Teflon scrim.

PULSE-JET FABRIC FILTERS FOR THE U.S. ELECTRIC UTILITY INDUSTRY

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ABSTRACT

Interest in pulse-jet fabric filters by electric utility companies in the U.S. is increasing rapidly. The interest stems from the commitment of utilities in Australia, principally the Electricity Commission of New South Wales, who have replaced, or are replacing, nearly 2000 MW of underperforming electrostatic precipitators (ESPs) with pulse jets. This significant retrofit program has attracted the attention of several U.S. utility companies who are considering pulse-jet fabric filters as a retrofit solution for poorly performing hot-and cold-side ESPs, for combined SO2 and particulate control, and as a compliance remedy in response to the recently promulgated PM10 ambient standard for fine particulates.

To obtain particulate emission control and pressure drop performance data for pulse jets collecting fly ash from coal-fired utility boilers, the Electric Power Research Institute (EPRI) has entered into a jointly funded R&D project with the Electricity Commission of New South Wales (ECNSW) in Australia. The project is being conducted at the Eraring station of the ECNSW to capitalize on the availability of two 5000 acfm (1 MW) pulse-jet pilot plants. The objectives are to conduct screening characterization tests on three generic types of pulse cleaning systems. The results will allow EPRI and the U.S. utility industry to assess the potential of pulse jets for particulate control in the U.S., and will provide the ECNSW with data for use in future procurement activities. The purpose of this paper is to summarize the factors that merit consideration of pulse-jet technology by U.S. utility companies, and to present preliminary data from the Australian pilot plant program.

INTRODUCTION

Fabric filters (baghouses) are firmly established in the U.S. for high-efficiency particulate control on utility coal-fired power plants. Since 1973 the use of fabric filters increased from zero to over 24,000 MW equivalent capacity (1). The fabric filters installed were exclusively reverse-gas and shake/deflate designs, and were constructed in sizes up to 850 MW. Although pulse-jet technology was available during this period, it was not applied to utility installations because of concerns regarding the mechanical complexity of the type of pulse jets that were available - primarily high-pressure, off-line cleaned units that used woven fiberglass bags. Further, these units did not deliver the bag life and level of emissions control needed by utility companies. Today, pulse jets are extremely competitive with more conservatively designed reverse-gas and shake/deflate baghouses due to two significant technological developments: (1) lower-pressure, on-line pulse cleaning, and (2) improved filter bag designs and materials. Consequently, pulse jets, which operate at sizes 40-50% smaller and at 30-40% lower cost than conventional reverse-gas and shake/deflate fabric filters, can achieve comparable reliability and particulate matter collection efficiency.

DISTINGUISHING FEATURES OF PULSE-JET FABRIC FILTERS

Pulse jets have many distinguishing design and operating features compared to reverse-gas and shake/deflate fabric filters. These features fall into the categories of design, filtration principles, and fabrics. Each category is discussed below.

DESIGN

Perhaps the most obvious distinction between pulse-jet and reverse-gas fabric filters is the direction of gas flow through the bags. In pulse jets the dirty gas flows from the outside of the bag to the inside, with the particulate matter collected on the outside of the bag. As shown in Figure 1, the bags are hung from the tubesheet, are closed at the bottom, and are open at the top where the clean gas exits the bag. Further, to support the bag during filtering pulse jets require an internal wire cage within the bag. Since the dust is collected on the outside of the bags, pulse-jet fabric filters are commonly called "outside collectors." In contrast, with reverse-gas and shake/deflate fabric filters flue gas flows in the opposite direction, that is, from the inside to the outside of the bag. The bags are normally open at the bottom where the dirty gas enters the bags, and they are closed at the top. Thus, the particulate matter is deposited on the inside of the bag, leading to the common description "inside collector."

Another significant difference between pulse-jet fabric filters and inside collectors is the methods used to clean the bags. In general, pulse jets use a short, powerful burst of air that is introduced at the top of the bag in the reverse direction to filtration as illustrated in Figure 1. The air burst propagates rapidly as a pressure wave down the length of the bag, causing the bag to accelerate away from the cage and then to immediately collapse back on the cage. This rapid flexing of the fabric dislodges the dust, which falls into the ash collection hopper below.

All baghouse types are compartmentalized, but only pulse jets are typically designed to clean the bags on-line without isolating compartments from the flue gas flow. This is accomplished by sequentially pulsing selected bags while the remainder continue to filter flue gas. On-line cleaning greatly simplifies overall fabric filter operation. However, since some of the dislodged dust is immediately reentrained back onto the bags, it may be necessary to use off-line cleaning for certain applications to maintain pressure drop and outlet emissions within acceptable bounds. On the other hand, due to the low-energy cleaning methods used with inside-collectors, bags must always be isolated from the flue gas flow to achieve effective cleaning (off-line cleaning). In either case, the compartmentalized design permits maintenance to be performed without requiring an outage of the entire fabric filter.

There are approximately 60 vendors of pulse-jet fabric filters, and the design and operation of the pulse cleaning systems vary widely (2). However, as illustrated in Figure 2 pulse cleaning can be categorized into three general regimes: high-pressure/low-volume, intermediate-pressure and volume, and low-pressure/high-volume pulsing. Traditionally, pulse-jet suppliers in the U.S. have used high-pressure/low-volume pulsing with woven fiberglass bags and off-line cleaning. In this approach, a high pressure (90 psig) burst of air is injected through an array of nozzles (blow pipes) located above the bags, which are arranged in a rectangular configuration. A venturi is positioned at the top of each bag to induce secondary air flow (actually cleaned flue gas) to assist in bag cleaning. High-pressure pulsing has a relatively high power consumption and, since the pulse dissipates rapidly as it travels down the bag, bag length is frequently limited to less than 12 feet.

Both intermediate-pressure/intermediate-volume and low-pressure/high-volume pulsing are relatively new developments in pulse cleaning. Intermediate-pressure cleaning also utilizes stationary nozzles above each bag similar to high-pressure cleaning. However, a pulse pressure of approximately 30 psig is used, and venturis are not needed above the bags to induce secondary air to assist in cleaning. Low-pressure/high-volume pulsing utilizes a novel, circular arrangement of oval-shaped bags, with nozzles attached to a rotating cleaning manifold located above the bags. In this design, a higher volume pulse of only 10 psig is used, which permits the use of blowers rather than compressors to supply the pulse cleaning air. Distinctive features of these "lower-pressure" pulse cleaning systems include: (1) bag length up to 20 ft., (2) reduced cleaning energy, (3) fewer pulse distribution valves, (4) on-line cleaning, and (5) simplified configurations and components. Both of these commercially available pulse cleaning systems have proven to be quite effective in removing dust from the bags (3-7).

By comparison, bag cleaning for inside-collectors is accomplished either by reverse-gas or a combination of mechanical shake with a small quantity of reverse gas "deflation" flow (8). In the former, clean flue gas is introduced opposite the filtering direction to collapse the bags and dislodge the dust, which falls into the ash collection hopper below. Anticollapse rings are sewn into the bag to prevent complete collapse of the bag during cleaning, thus allowing the dislodged ash to fall unobstructed into the hopper. In shake/deflate cleaning, deflation gas (consisting of clean flue gas) is introduced in the opposite direction to filtration to overcome damper leakage and to relax the bags prior to shaking. Essentially all the dust removal is accomplished by the shaking action. "Shaker" cleaning is a variation of this method that is used in Australia, and consists of vigorously shaking the bags without deflation gas.

The more energetic cleaning of pulse-jet fabric filters, coupled with more frequent cleaning, results in lower resistance to gas flow than can be achieved with inside collectors. This lower resistance, defined in terms of pressure drop through the unit, makes it possible to operate pulse jets at a much higher air-to-cloth ratio (A/C), which is the ratio of flue gas volume flow (acfm) divided by total fabric filtering area (ft^2). For fly ash applications pulse-jet fabric filters are typically twice the A/C, or approximately one half the size, of inside-collector baghouses. Figure 3 compares the plan area required for reverse-gas and pulse-jet fabric filters installed on a 350 MW pulverized-coal-fired utility boiler. The pulse jet used in this example is a low-pressure/high-volume unit with a circular bag configuration, and the dimensions for both units are based on actual design information. As the figure shows, the pulse jet requires approximately 45% less plan area than the reverse-gas unit.

FILTRATION PRINCIPLES

Another significant difference between pulse-jet and inside-collector fabric filters is the principles governing filtration. Two distinctive modes arise from the cleaning methods and types of fabrics used. With inside-collectors, low-energy cleaning and woven fabrics lead to the establishment of a permanent dustcake on the bags that tends to increase with time. The dustcake is the primary filtering medium, whereas the role of the bag is to provide a support structure for the dustcake. On seasoned bags dustcake weights typically achieve levels of 0.6 to 2.0 lb/ft^2 for reverse-gas fabric filters, and 0.3 to 0.6 lb/ft^2 for shake/deflate units (8). With pulse-jet fabric filters the use of high-energy cleaning minimizes the formation of a permanent dustcake, and residual dustcake weights on seasoned bags are generally less than 0.2 lb/ft^2. Thus, the bag fabric acts as the primary filtering medium, and must be constructed with the appropriate permeability to yield high collection efficiency and low pressure drop.

Although the filtration mechanisms for pulse-jet and inside-collectors are quite different, the operating goals are similar: high particulate collection efficiency (99.9%), moderate pressure drop (5 in. H2O), and good bag life (3 years or better). The filtration principles described above, coupled with the dust properties and bag type used, dictate the operating conditions required for each type of collector to achieve these goals. For fly ash filtration, air-to-cloth ratios less than 2.5 acfm/ft^2 are generally required for inside collectors, whereas air-to-cloth ratios less than 4.5 acfm/ft^2 are typical for pulse jets.

FABRICS

Considering the more energetic cleaning action and high air-to-cloth ratio associated with pulse jets, felted fabrics are generally preferred over woven materials to achieve high fly ash collection efficiency and low pressure drop. When pulse-jet fabric filters where first applied to coal-fired boilers, Nomex and Teflon were the only fibers capable of continuous operation above 275°F (9). However, Nomex proved unsuitable because of its susceptibility to hydrolysis under the hot, humid, and acidic flue gas conditions characteristic of coal-fired boilers. Chemical treatments or finishes have since been developed to improve the performance of Nomex under these conditions, and good results have been reported for pulse-jet installations overseas. Teflon felt has also been used successfully in coal-fired boilers overseas, although its high cost and dimensional stability problems experienced at temperatures above 400°F may limit applications in the U.S. Fiberglass fabrics were developed to meet the needs for a low-cost fabric suitable for operation at coal-fired boiler outlet temperatures (300°F). Fiberglass fabrics are strong, can withstand high flue gas temperatures (i.e., up to 500°F), and can be texturized to control dustcake formation. Pulse-jet experience with fiberglass has been principally in high-pressure/low-volume units using off-line cleaning. The bulk of this experience has been with woven fabrics, although experimental glass felt bags (Huyglas) have been used in some installations. Despite extensive use of fiberglass fabrics in pulse jets filtering fly ash from coal-fired boilers, the results have been inconsistent. For example, bag lives of three to four years have been achieved in some units, whereas less than 18 months life has been experienced in many others. Poor life with fiberglass bags is frequently attributed to abrasive damage of the bags by the support cages due to improper fit or corroded cages, and to acid attack of the fabric (9). Although good performance can be achieved with fiberglass, special precautions are necessary: (1) the fabric must be properly constructed and finished for the specific application, (2) cage design, bag-cage fit, and proper bag installation procedures are critical, and (3) off-line cleaning is necessary, particularly for woven glass bags.

More recently, felts such as Dralon T (homopolymer acrylic) and Ryton (polyphenylene sulfide) have been developed and used successfully in coal-fired boiler applications. Acrylic felts offer superior mechanical resilience to fiberglass at comparable cost, and have achieved good success in Australian pulse-jet applications. The primary disadvantage of acrylic is the 270°F maximum continuous operating temperature specified for the fabric. This is below the outlet temperature of many U.S. utility boilers, which range from 225°F to 280°F for low-sulfur coals, and 260°F to 350°F for high-sulfur coals. Although acrylic has been used successfully in Australia for low-sulfur coal on both shaker and pulse-jet fabric filters, standby backend cooling systems are employed as a protective measure for the bags in the unlikely event that a flue gas temperature excursion above the acrylic operating range occurs.

In the United States, acrylic fabrics are a worthwhile option for utility companies to consider for pulse jets operated in conjunction with spray-drying and dry SO_2 removal processes where flue gas temperatures will be well below 270°F.

Ryton felts also offer outstanding abrasion resistance at a cost of 1.7-3 times fiberglass, and are suitable in moist, acidic environments up to 400°F. Although Ryton was not available in commercial quantities until 1984, it has been demonstrated to be an excellent fiber for pulse-jet fabric filters . The initial use of Ryton was for Daytex felted bags developed by the Carter-Day Company in the late 1970s. Daytex, which consists of Ryton felt with a Rastex (Gore-Tex PTFE fiber) scrim, has achieved bag lives exceeding seven years in pulse jets installed on industrial pulverized-coal-fired boilers at Coors (3,4). More recently, Ryton felt bags with a woven Ryton scrim have been developed and are being operated successfully in pulse-jet baghouses overseas.

In summary, three fabric options appear to be most suitable for pulse-jet fabric filters in the U.S. utility industry on the basis of cost and reliability: Ryton/Ryton felt, Daytex felt, and woven or felted fiberglass. Dralon T (acrylic) is a worthwhile alternative for low temperature situations. Research and development activities currently being supported by equipment and fabric suppliers are seeking to develop a superior woven fiberglass bag for pulse jets that will achieve comparable performance to Ryton and acrylic. Furthermore, R&D activities are being supported by these same groups to reduce the cost of Ryton bags, and to improve the integrity of Nomex bags. Table 1 summarizes some important characteristics of bag fabrics generally available for pulse-jet fabric filters collecting fly ash in the electric utility industry.

PULSE-JET FABRIC FILTER PERFORMANCE

In addition to the bag life considerations discussed above, particulate emissions control, pressure drop, and economics are major selection criteria for U.S. utility companies contemplating pulse-jet fabric filters. Discussions of these criteria are presented below.

PARTICULATE EMISSIONS

Figure 4 compares outlet particulate emissions for reverse-gas and shake/deflate baghouses using woven fiberglass bags, and low-and intermediate-pressure, pulse-jet fabric filters using Daytex, acrylic, and fiberglass felted bags. The data are based on measurements at pilot-and full-scale fabric filters collecting fly ash from both utility and industrial pulverized-coal-fired boilers (3-8, 10-15). Moreover, the pulse-jet data are for units using on-line cleaning. The white band indicates the range of emissions that would be expected for commercial units. The bottom of the white band is indicative of data from pilot units and full-scale fabric filters in good operating condition (i.e. relatively new bags), whereas the top of the band is more representative of emissions for units with several years of service on a given set of bags. The data show that outlet particulate emissions increase as more energetic bag cleaning methods are used, and that the emissions from shake/deflate and pulse-jet fabric filters are approximately equivalent. However, for all the fabric filter types outlet emissions are below the 1979 New Source Performance Standards (NSPS) of 0.03 lb/10^6 Btu. Furthermore, outlet emissions within all the ranges shown in Figure 4 achieve a clear stack plume.

PRESSURE DROP

Figure 5 compares pressure drop performance for reverse-gas and low-and intermediate-pressure, pulse-jet fabric filters for typical operating conditions on coal-fired boilers (3-8, 10, 12, 13, 15). As the figure illustrates, pulse jets can be operated at twice the air-to-cloth ratio of reverse-gas units while maintaining comparable pressure drop. Figure 6 shows a similar comparison between shake/deflate and pulse-jet fabric filters (8, 11, 13, 14). Again, pulse jets retain a significant performance advantage, which results in an air-to-cloth ratio that is 1.5 to 2 times higher than for shake/deflate fabric filters at a given pressure drop.

ECONOMICS

Figures 7 and 8 present capital cost comparisons of electrostatic precipitators (ESPs), reverse-gas fabric filters, and low-pressure, pulse-jet fabric filters for both Powder River Basin and Appalachian coals (16, 17). EPRI economic premises were used in the analysis (18), which shows that for low-sulfur, Powder River Basin coal, new unit capital costs for pulse-jet fabric filters are estimated to be $47/kW. In comparison, the estimated capital costs for reverse-gas fabric filters and electrostatic precipitators are $68/kW and $86/kW, respectively. For Appalachian 2% sulfur coal, new unit capital costs for pulse-jet fabric filters are estimated to be $40/kW, whereas for reverse-gas fabric filters and electrostatic precipitators capital costs are projected to be $57/kW and $55/kW, respectively. Although not shown here, 30 year levelized costs show similar relationships among the three collector types. Retrofit costs would in all probability be higher for each control technology, although these costs are difficult to estimate without site-specific design information. Nevertheless, it is expected that the cost differentials for the three technologies would be retained in most retrofit situations.

AUSTRALIAN PILOT PLANT R&D PROGRAM

To confirm the results reported above, the Electric Power Research Institute (EPRI) and the Electricity Commission of New South Wales (ECNSW) in Australia are jointly sponsoring an R&D project to quantify the particulate emission control and pressure drop performance of low-pressure (10 psig), intermediate-pressure (30 psig), and high-pressure (90 psig) pulse-jet fabric filters. The data collected will allow EPRI and the U.S. utility industry to assess the potential of pulse jets for particulate control on coal-fired utility boilers, and to determine the need for a pulse-jet R&D program in the United States. Moreover, the ECNSW will obtain valuable data for use in future procurement activities.

The project is being conducted at the Eraring station of the ECNSW, which burns an Australian coal that is approximately 0.4% sulfur and 25% ash. The benefits of performing the project at Eraring include the availability of two 5000 acfm (1 MW equivalent) pulse-jet pilot plants that can be operated independently to evaluate the range of pulse pressures described above. One plant is a low-pressure/high volume (LPHV) unit of similar geometry to one module of a full-scale pulse jet as shown in Figure 9. The pilot unit contains 48 full-sized bags arranged concentrically in three rows. The second unit simulates both high-pressure/low-volume (HPLV) and intermediate-pressure/intermediate-volume (IPIV) pulse cleaning modes. The pilot unit is similar to the schematic diagram shown in Figure 1, and contains three rows of bags. For IPIV operation each row contains 18 bags (54 bags total), whereas for HPLV operation each row has 12 bags (36 bags total). Although this pilot unit is not intended to represent any particular manufacturer's product, it is suitable for conducting R&D studies of high- and intermediate-pressure pulse cleaning. A photograph of the pilot plant test facility is shown in Figure 10.

The pilot study at Eraring is scheduled to continue for two years. A variety of operating conditions will be evaluated in the LPHV pilot plant throughout the two year project. In the second pilot unit, intermediate-pressure/intermediate-volume pulsing will be tested for approximately 12 to 16 months. If budget permits, the pilot unit will then be modified for high-pressure/low-volume pulse cleaning, and a 2000 hour test program will be conducted. The ECNSW is monitoring the test program and conducting particulate emission tests. Southern Research Institute is acting on behalf of EPRI to monitor the program with the ECNSW, to review the test data, and to perform supporting laboratory analyses. The data collected include pressure drop, boiler load, flue gas volume flow rate, inlet and outlet particulate matter concentrations, fly ash size distribution, outlet opacity, coal and ash analysis, flue gas temperature, pulse cleaning parameters, bag weights, and bag integrity information.

The LPHV and IPIV pilot units are equipped with 20 ft. long Daytex felted bags, are operated at a constant gas volume flow rate (air-to-cloth ratio) of 4 acfm/ft^2, and are cleaned on-line. The LPHV plant is cleaned with air supplied at 10 psig, and the IPIV plant is cleaned with air supplied at 30-50 psig. The primary data being used to compare the operation of the two pilot plants are (1) pulse cleaning air consumption to maintain pressure drop within a specified operating range, and (2) particulate matter collection efficiency, outlet emissions, and outlet opacity. To simplify comparison of the pulse air cleaning requirements, bag cleaning for both pilot plants is initiated at a tubesheet pressure drop of 5 in. H2O and continues until the pressure drop decreases to approximately 4.5 in. H2O. Pulse frequency, pulse count, and pulse tank pressure are all continuously recorded for calculation of cleaning air consumption. Particulate emission data are collected approximately once a month, and outlet opacity is determined from a Triboflow continuous emission measurement device.

RESULTS

The LPHV and IPIV pilot plants were started up in July 1987 and have accumulated approximately 3000 hours of operation. Both units have maintained pressure drop between 4.5 and 5 in. H2O at an air-to-cloth ratio of 4 $acfm/ft^2$. Due to operating problems experienced with the IPIV plant, particulate matter emissions data are only available for the LPHV plant. Particulate matter collection efficiency measured at 1600 hours operation for the LPHV plant was 99.92%, with outlet emissions of 0.007 gr/dscf, or approximately 0.013 $lb/10^6$ Btu. Measurements conducted more recently at 3000 hours of operation showed outlet emissions of 0.01 gr/dscf, or approximately 0.018 $lb/10^6$ Btu. Accordingly, after 3000 hours operation, particulate matter emissions and pressure drop for the LPHV pulse-jet pilot plant are within the data ranges illustrated in Figures 4-6.

SUMMARY AND CONCLUSIONS

Pulse-jet fabric filters offer the U.S. electric utility industry a particulate emissions control technology that is applicable to both new and retrofit units. In particular, pulse jets are an attractive alternative for utility companies requiring retrofit solutions for poorly performing hot-and cold-side electrostatic precipitators, and as a compliance remedy in response to the recently promulgated PM 10 ambient standard for fine particulates. Due to the wide range of fabrics available and the energetic cleaning used, pulse jets promise to be a forgiving particulate control technology capable of delivering consistent performance for a variety of fuel sources and combustion systems. Furthermore, pulse jets are attractive because they can be factory assembled in modules, which greatly reduces cost and field erection time.

Evaluation of available performance data for pulse-jet fabric filters operating on pulverized-coal-fired boilers indicates that, with proper selection of fabrics and pulse cleaning systems, pulse jets can be operated at sizes 50% smaller than conventional reverse-gas fabric filters, yet still achieve high particulate collection efficiency.

Consequently, pulse jets can now compete directly with electrostatic precipitators and conventional reverse-gas fabric filters for particulate control on fossil-fuel-fired utility boilers, but at significant savings in cost. In view of these developments, it appears that pulse jets will be the next generation fabric filter technology for the U.S. utility industry. However, the majority of experience on coal-fired utility boilers is nondomestic. Therefore, it is appropriate for pulse jets to be demonstrated on U.S. coals before utility companies will have technical justification for making larger-scale commitments to the technology. The pilot R&D program currently being conducted in Australia will allow EPRI and the U.S. utility industry to assess the potential of pulse jets for particulate control on coal-fired utility boilers, and also to determine the need for a pulse-jet R&D program in the United States.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

ACKNOWLEDGEMENTS

The Australian pilot plant R&D program described in this paper is being supported by the Electricity Commission of New South Wales in Australia, and the Electric Power Research Institute under RP 1129. The authors gratefully acknowledge the contributions of Mr. Gordon Floyd and Mr. Col Robertson of the ECNSW for their willingness to participate in this cooperative R&D program, and also for providing the host site and testing support services. Furthermore, we acknowledge the contributions of EPRI project manager Mr. Walter Piulle, who organized the project for EPRI and helped develop the test plan. We also acknowledge the support of EPRI project manager Mr. Ramsay Chang, who has directed the project for EPRI since October 1987. Finally, we express our thanks to Mr. Raffi Andonian and Mr. Bruce Russell of Howden Environmental Systems for their tireless efforts in collecting the pulse-jet data and interpreting the results.

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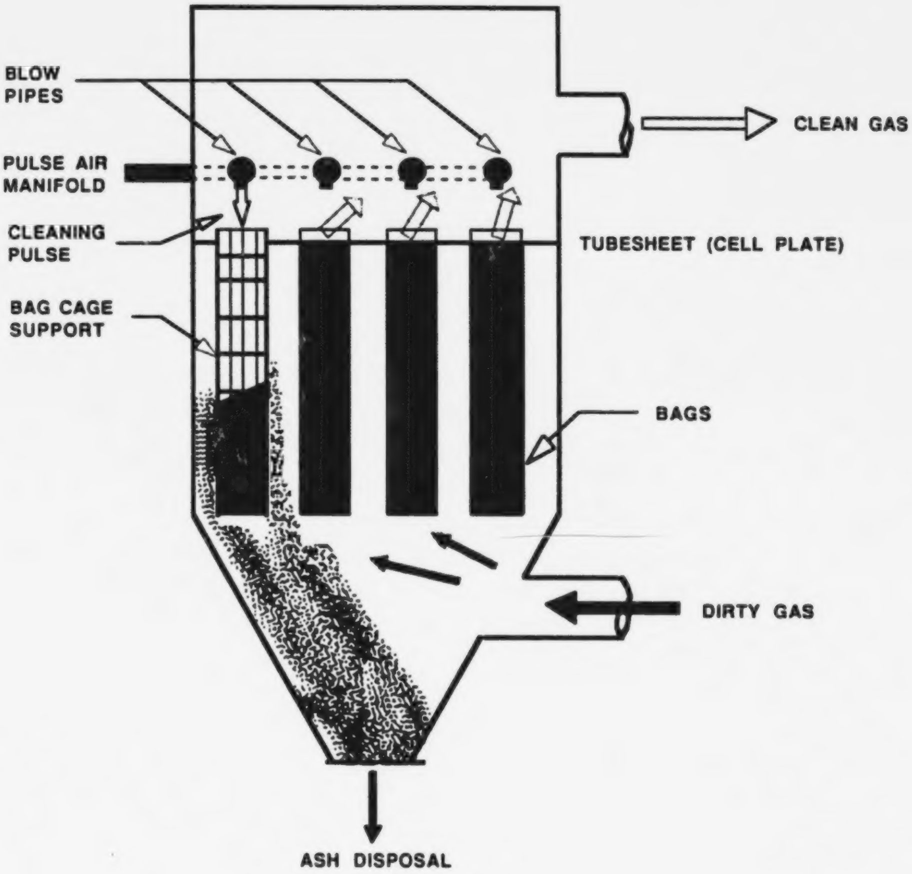
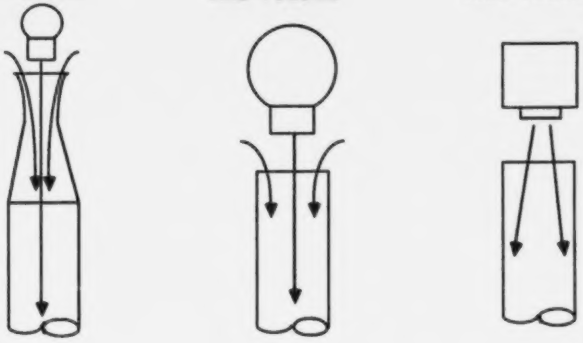
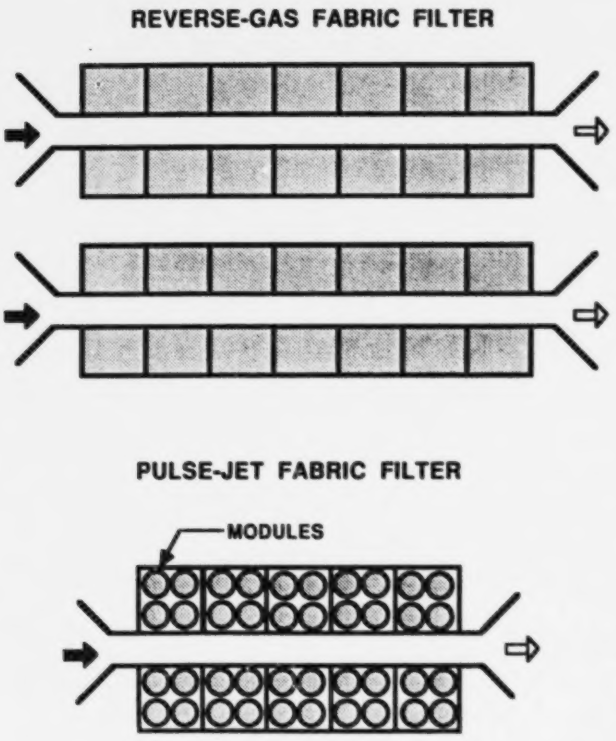


Figure 1. Schematic diagram of a typical pulse-jet fabric filter illustrating the primary components. The bags are fitted over support cages and are hung from the tubesheet. The pulse cleaning air is supplied either by compressors or blowers, and is distributed via a pulse air manifold system to blowpipes located over each bag. Cleaning is accomplished by sequentially pulsing selected bags in the compartment. The bags in the left row are being cleaned on-line, while the remaining bags in the compartment are filtering dirty gas.



TYPICAL CHARACTERISTICS	HIGH PRESSURE, LOW VOLUME	INTERMEDIATE PRESSURE AND VOLUME	LOW PRESSURE, HIGH VOLUME
TANK PRESSURE, PSIG	90	20 - 30	10
SECONDARY/PRIMARY AIR	6 - 7	1 - 2	- - -
BAG LENGTH, FT (ON-LINE CLEAN)	12	20	20
BAG DIAMETER, IN	4.5	5	6 X 2.7 (OVAL)
NO. BAGS/VALVE	12	18	484
CLOTH AREA/VALVE, FT^2	172	462	12,600
CLEANING ENERGY, W/FT^2	4	0.8	0.8

Figure 2. Comparison of typical operating characteristics for high-pressure, intermediate-pressure, and low-pressure pulse cleaning. Note the venturi required for high-pressure pulsing to induce secondary air (cleaned flue gas) to assist in cleaning. For high-pressure pulsing, woven fiberglass bags up to 20 ft. long have been used with off-line cleaning.



REVERSE-GAS FABRIC FILTER

A/C = 1.7
WOVEN GLASS 34 FT. BAGS
28 COMPARTMENTS
306 BAGS/COMPARTMENT
216' L x 126' W x 90' H
PLAN AREA: 25,320 FT^2

PULSE-JET FABRIC FILTER

A/C= 3.5
RYTON/RYTON 20 FT. BAGS
10 COMPARTMENTS
4 MODULES/COMPARTMENT
376 BAGS/MODULE
162' L x 89' W x 79' H
PLAN AREA: 13,608 FT^2

Figure 3. Comparison of plan arrangements for reverse-gas and pulse-jet fabric filters for a 350 MW pulverized-coal-fired utility boiler. The pulse-jet fabric filter is a low-pressure/high-volume unit with a circular arrangement of oval-shaped bags.

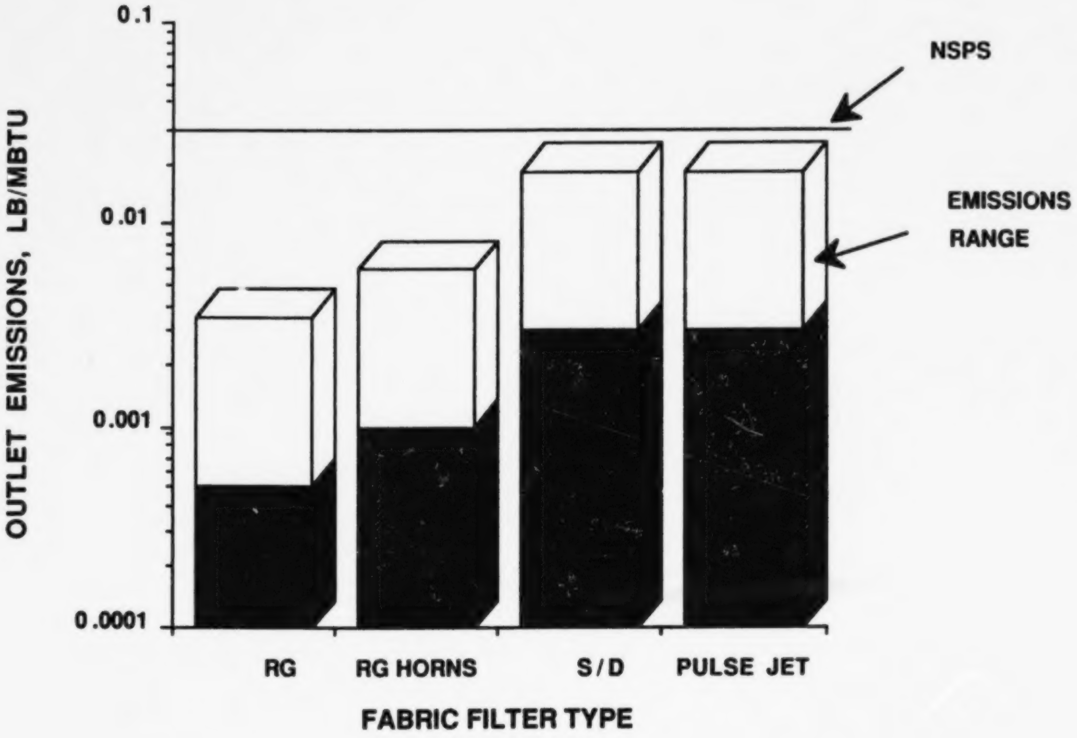


Figure 4. Comparison of outlet emissions for reverse-gas, reverse-gas with sonic assist, shake/deflate, and pulse-jet fabric filters. The white band is indicative of the emissions range expected for fabric filters operating in commercial practice. The pulse-jet data are for low-and intermediate-pressure units using on-line cleaning.

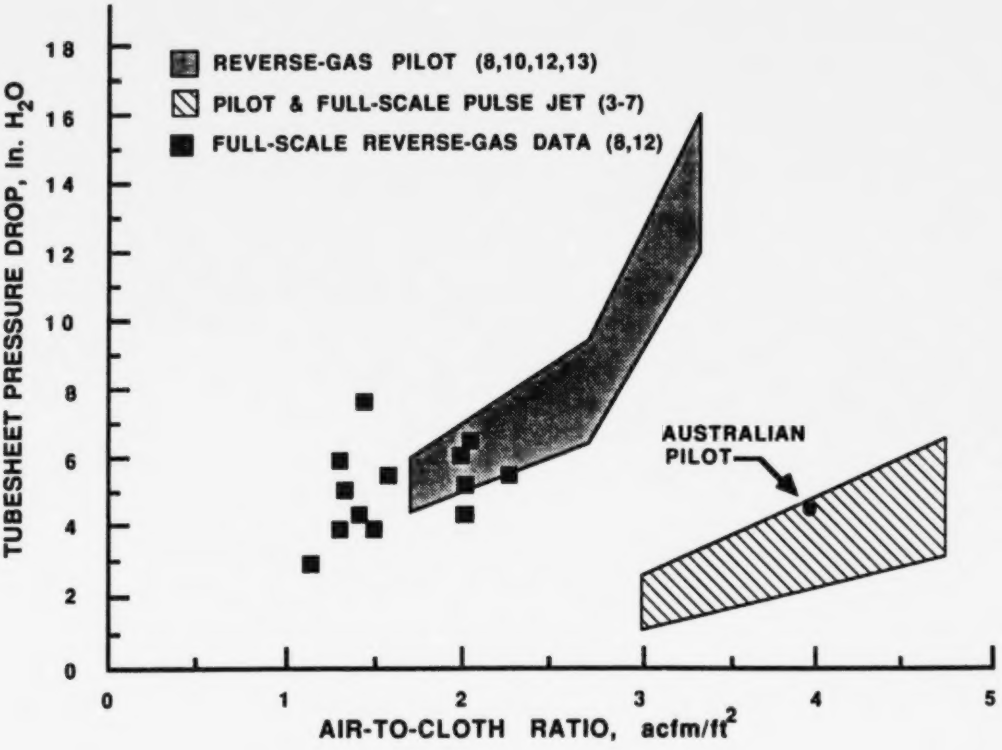


Figure 5. Comparison of pressure drop vs. air-to-cloth ratio for reverse-gas and pulse-jet fabric filters. The desirable operating range is in the lower right quadrant. The shaded area for pulse jets reflects the range of pilot- and full-scale data for low-and intermediate-pressure pulsing. The circular data point represents recent results from a jointly funded ECNSW/EPRI pulse-jet pilot study.

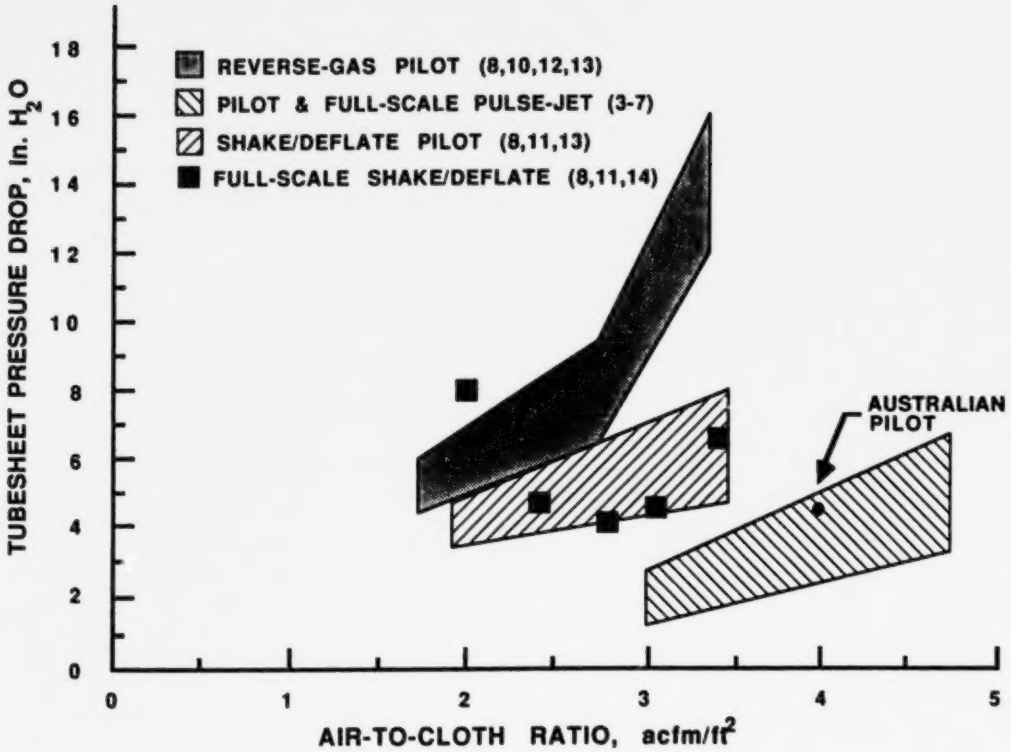


Figure 6. Comparison of pressure drop vs. air-to-cloth ratio for shake/deflate and pulse-jet fabric filters. The desirable operating range is in the lower right quadrant. The shaded area for pulse jets reflects the range of pilot- and full-scale data for low-and intermediate-pressure pulsing. The circular data point represents recent results from a jointly funded ECNSW/EPRI pulse-jet pilot study.

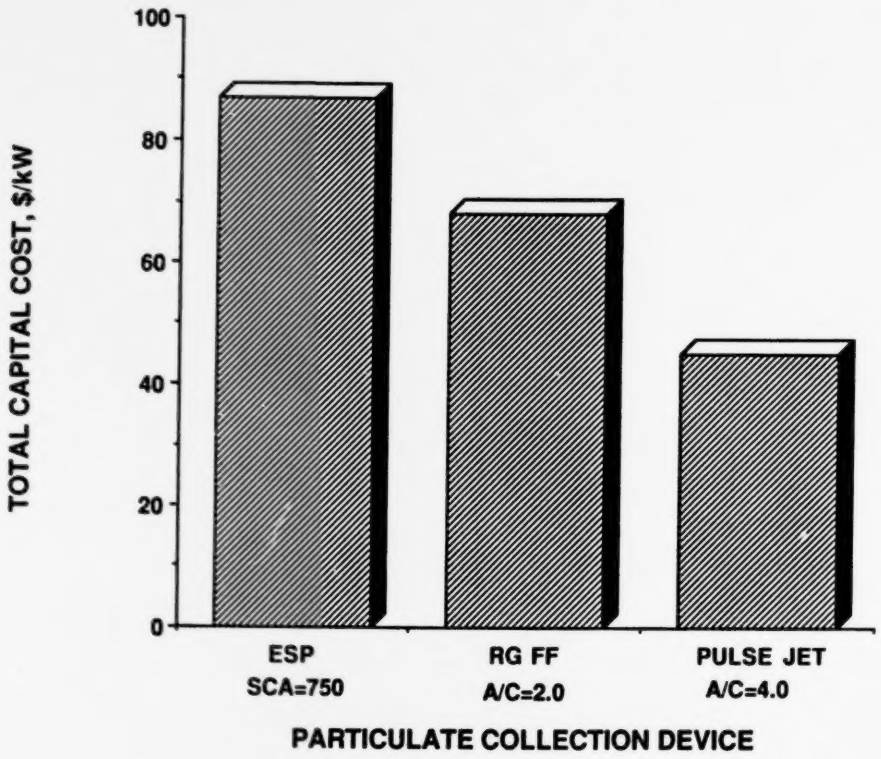


Figure 7. Comparison of capital costs for an electrostatic precipitator, reverse-gas fabric filter, and a low-pressure/high-volume, pulse-jet fabric filter. The results are based on a 350 MW pulverized-coal-fired boiler meeting New Source Performance Standards for particulate matter of 0.03 lb/10^6 Btu, and firing a low-sulfur, Powder River Basin coal.

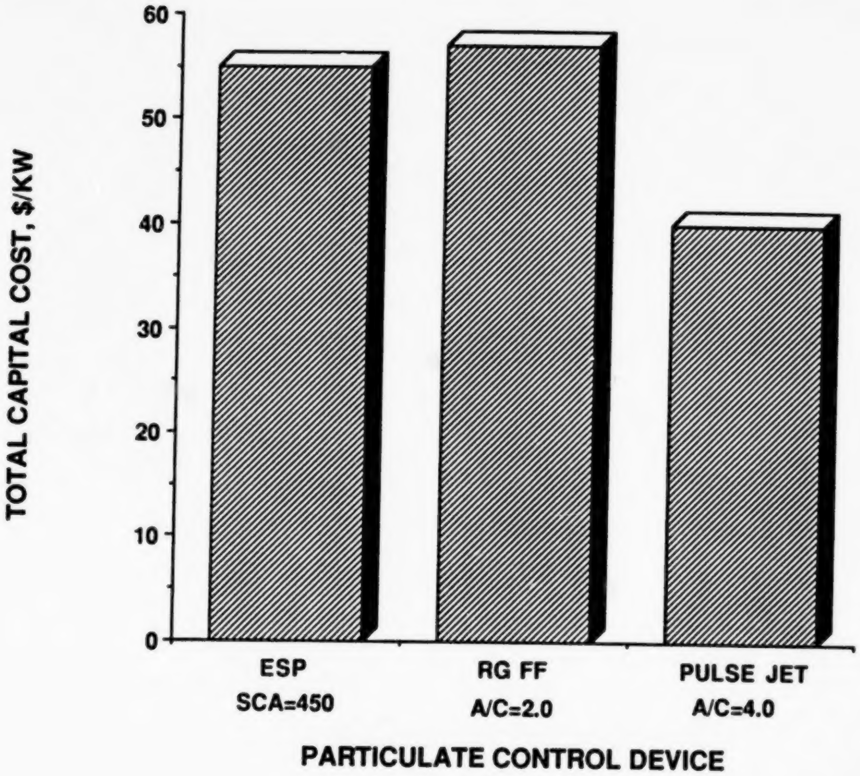


Figure 8. Comparison of capital costs for an electrostatic precipitator, reverse-gas fabric filter, and a low-pressure/high-volume, pulse-jet fabric filter. The results are based on a 350 MW pulverized-coal-fired boiler meeting New Source Performance Standards for particulate matter of 0.03 lb/10^6 Btu, and firing an Appalachian, 2% sulfur coal.

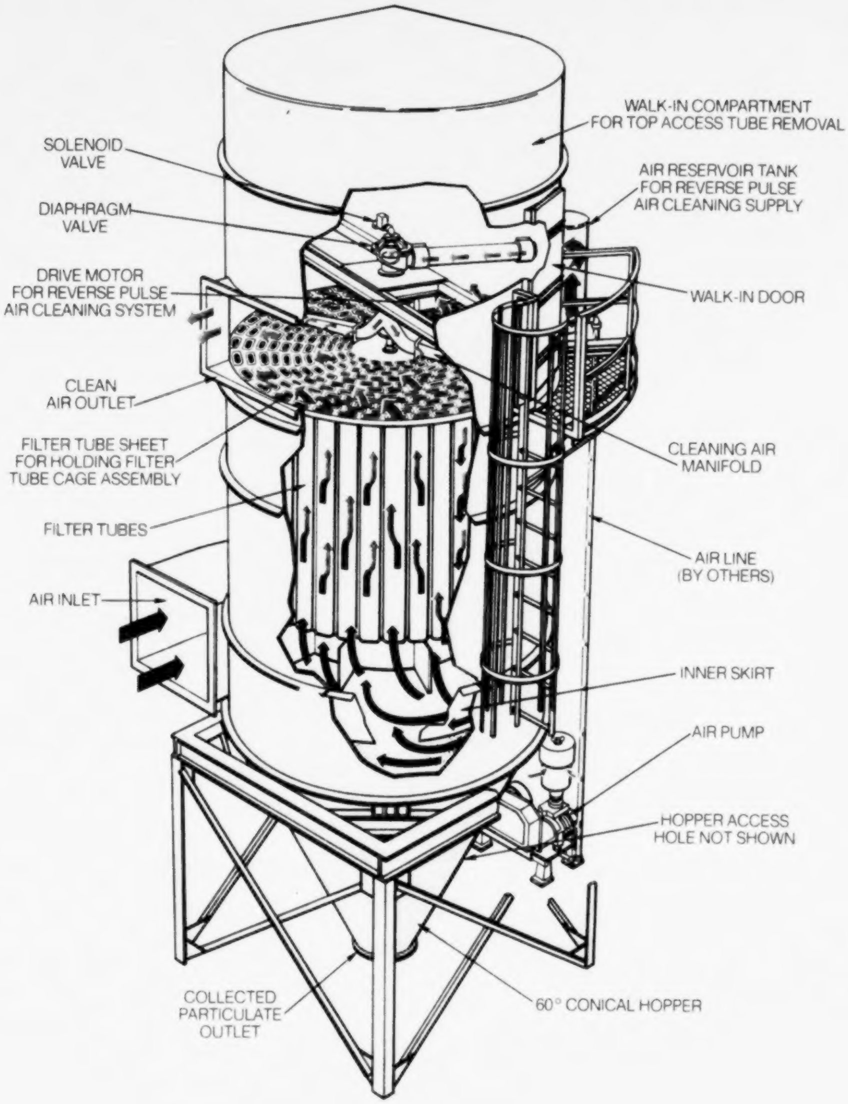

WALK-IN COMPARTMENT FOR TOP ACCESS TUBE REMOVAL
SOLENOID VALVE
AIR RESERVOIR TANK FOR REVERSE PULSE AIR CLEANING SUPPLY
DIAPHRAGM VALVE
DRIVE MOTOR FOR REVERSE PULSE AIR CLEANING SYSTEM
WALK-IN DOOR
CLEAN AIR OUTLET
FILTER TUBE SHEET FOR HOLDING FILTER TUBE CAGE ASSEMBLY
CLEANING AIR MANIFOLD
FILTER TUBES
AIR LINE (BY OTHERS)
AIR INLET
INNER SKIRT
AIR PUMP
HOPPER ACCESS HOLE NOT SHOWN
COLLECTED PARTICULATE OUTLET
60° CONICAL HOPPER

Figure 9. Schematic of one compartment of a low-pressure/high-volume, pulse-jet fabric filter supplied by Howden Environmental Systems. Oval bags are arranged in concentric circles, with up to 484 bags per compartment. The bags are cleaned by pulses from the manifold arm which rotates at one revolution per minute. Cleaning air, supplied to the manifold by blowers, is introduced to the manifold through a central dump valve. The cleaning air is injected into the bags via nozzles in the pulse air manifold.

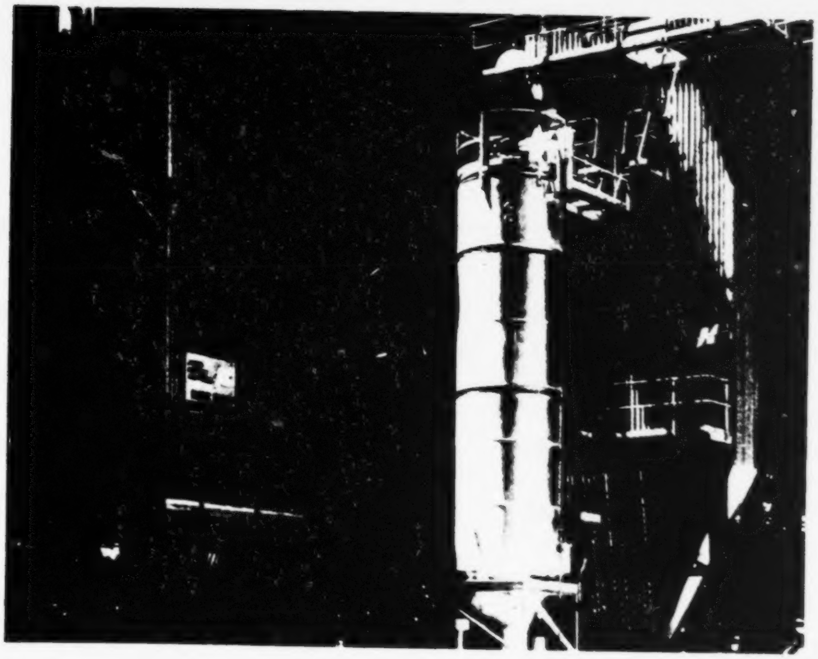


Figure 10. Fabric filter pilot plant test facility located at the Eraring station of the Electricity Commission of New South Wales in Australia. The low-pressure pulse jet is in the center, whereas the intermediate-pressure unit is on the right. On the left is a shaker cleaned baghouse pilot plant. All the pilot units are designed to filter approximately 5000 acfm (1 MW equivalent) of flue gas.

FABRIC	MAXIMUM TEMPERATURE, °F	ABRASION/ACID RESISTANCE	RELATIVE COST
WOVEN FIBERGLASS	500	POOR/GOOD	1
FELTED NOMEX	425	GOOD/POOR	1
FELTED ACRYLIC	285	VERY GOOD/GOOD	0.8
FELTED DAYTEX	400	VERY GOOD/GOOD	3
FELTED RYTON	400	VERY GOOD/GOOD	1.7
FELTED TEFLON	475	FAIR/EXCELLENT	10

Table 1. Characteristics of fabrics generally available for pulse-jet fabric filters used to collect fly ash from coal-fired utility boilers.

MECHANISMS OF FABRIC FILTER PERFORMANCE IMPROVEMENT WITH FLUE GAS CONDITIONING

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ABSTRACT

Particulate emissions from a fabric filter and tube sheet pressure drop can be significantly reduced by injecting small amounts of gas conditioning agents, ammonia and sulfur trioxide (SO_3), upstream of the filter. The conditioning agents make the ash particles more cohesive which reduces the seepage of dust through the fabric and facilitates the bridging of pinholes, preventing direct particle penetration. At the same time, a more porous dust cake is formed which results in reduced baghouse pressure drop.

Recent tests have been conducted to further investigate the mechanisms by which conditioning agents improve fabric filter performance. A review of penetration mechanisms shows that there is a theoretical basis for lower emissions with increased bulk cohesion. Pressure drop reduction as a result of conditioning is attributed to an increase in dust cake porosity as theoretical and empirical models predict.

INTRODUCTION

Since the purpose of a baghouse is to filter the flue gas of fine suspended fly ash, the first measure of performance must be the collection efficiency, or more accurately, the level of particulate emissions from the baghouse. New Source Performance Standards for utility coal-fired boilers limit particulate emissions to 0.03 lb/million Btu and require less than 20% opacity. The particulate removal efficiency required to meet this standard is about 99.5% for an inlet grain loading of 3 grains/scf. In general, baghouses can easily meet this requirement, and in some cases, particulate emissions from baghouses may be several orders of magnitude less than the 0.03 lb/million Btu standard. In spite of the low particulate emissions typically observed with baghouses, there are a number of reasons for improving the collection efficiency. Other aspects of the Clean Air Act may require higher levels of control, including visibility protection and prevention of significant deterioration provisions. Visibility protection is a major issue for plants located near designated Class I areas such as national parks, and prevention of significant deterioration may severely limit emissions when a plant is located in an area which is already near the allowable increment in emissions. State and local regulatory agencies may also impose stricter emission requirements. In addition, the new PM10 ambient air quality standard means that, in the future, plants may be scrutinized for fine particle emissions as opposed to total mass particulate emissions. Although reported particulate removal efficiencies for utility baghouses range from 99.0% to 99.99+% (1), high emissions and/or high opacity spikes after bag cleaning have also been reported, indicating that baghouse particulate emissions are fuel specific (2-5). Lab-scale and pilot-scale data show that some fly ashes can be much more difficult to collect in fabric filters than others (6,7,8). Therefore, a need exists to improve collection efficiency for those baghouses filtering difficult-to-collect ashes. There is also concern because fine fly ash particles are enriched in trace elements (9) that have highly toxic effects on humans and are suspected or known carcinogens. Furthermore, health studies have indicated that current ambient concentrations of fine particulate matter are a threat to public health (10).

Another important aspect of fabric filter performance is baghouse pressure drop for which control is usually improved by changing mechanical parameters such as bag cleaning, fabric type, or face velocity. An alternative method is to change the characteristics of the dust. Conditioning agents have long been applied to improve electrostatic precipitator performance by changing dust characteristics such as resistivity and bulk cohesion, but changing dust characteristics to enhance fabric filter performance is a relatively new concept. Ammonia addition has been tried on a test basis at the Monticello plant as a means of reducing particulate emissions (3). Previous work at the University of North Dakota Energy and Mineral Research Center has shown that significant reductions in fine particulate emissions and pressure drop can be achieved by the simultaneous use of ammonia and SO_3 as conditioning agents (11-15). While the previous work shows the potential for improving fabric filter performance (emissions and pressure drop), questions remained concerning the mechanisms of improvement. The purpose of this paper is to explore the mechanisms by which conditioning agents improve fabric filter performance.

EMISSIONS REDUCTION WITH FLUE GAS CONDITIONING

PILOT-SCALE CONDITIONING TESTS

Previously reported conditioning tests were conducted with a 550,000 Btu/hr pulverized-coal fired combustor and pilot baghouse (11-15). Ammonia and SO_3 were injected into the flue gas upstream of the baghouse at temperatures of 700°F and 350°F. Injection at both temperatures resulted in significant reductions in fine particulate emissions. For all tests, the baghouse temperature was 300°F. Concentrations for the first tests were 40 ppm ammonia and 12 ppm SO_3, but later tests showed that 25 ppm ammonia and 12 ppm SO_3 were adequate. The fabric for most of these tests was a 10 oz woven glass 3 x 1 weave with texturized fill yarns and 10 % Teflon B finish (MIDWESCO Filter Media 601 E). This cloth is typical of woven glass fabrics used in utility baghouses. The air-to-cloth ratio for most tests was 3.3 ft/min. All tests were started with new fabric. For the earlier tests, bag cleaning was accomplished by taking 3 of a total of 9 bags off line for shaker cleaning every 20 minutes. For later tests, all 9 bags were shaker cleaned off line every 2 hours. Typical tests were for a duration of 10 to 12 hours, however tests up to 50 hours were also completed. For 6 different coals, baseline and conditioning tests were conducted to determine the effect of conditioning on particulate emissions and pressure drop. Particulate emissions in the range of 0.5 to 15 μm were measured as a function of time with an APS 33 aerodynamic laser particle sizer. Total mass particulate emissions were determined via EPA Method 5, and submicron emissions were measured by a Differential Mobility Particle Sizer and/or condensation nucleus counter.

Typical emissions results from these conditioning tests using South Hallsville and Monticello coals are shown in Figures 1 and 2. Respirable mass as measured by the APS 33 is a sum of particle mass between 0.5 and 10 μm including all particles from 0.5 to 1.5 μm and a decreasing percentage of the particle mass from 1.5 to 10 μm. The reduction in respirable mass emissions is about 4 orders of magnitude for the South Hallsville coal and 3 orders of magnitude for the Monticello coal. The inlet respirable mass for these coals was about 1100 mg/m^3. This means that the respirable mass collection efficiency has improved from about 90% to 99.999% for South Hallsville and to about 99.99% for the Monticello coal. The dust loading and submicron emissions measurements also confirm this significant reduction in particulate emissions when using ammonia and SO_3 as conditioning agents. These two coals produce ashes that are difficult to collect in a fabric filter. However, tests with other coals that produce somewhat less difficult-to-collect ashes also resulted in some reduction in particulate emissions with conditioning.

BENCH-SCALE TESTS

Following evaluation of the conditioning results generated with the pilot combustor and baghouse, a bench-scale fabric filter test system was constructed to evaluate mechanisms of fabric filter performance improvement. A schematic of the system is shown in Figure 3. Major components of the system include the dry powder disperser, filter holder, and particulate measuring instrumentation. The system is equipped with an SO_3 and ammonia injection system and is designed so that the filter can be heated to temperatures typical of baghouses. However, bench-scale results reported in this paper include only tests conducted at ambient temperature and without additional conditioning agents. The gas flow rate through the system is designed to be 25 liters/min (0.88 scfm) at a filter face velocity of 2 ft/min under standard conditions. The filter holder is designed for a square fabric sample with a maximum filtering area of 0.415 ft^2. Fly ash is introduced to a cylindrical chamber in the TSI Model 3410 Dry Powder Disperser. A piston then moves into the chamber at an adjustable, uniform speed pushing the fly ash against a stainless steel brush rotating at high speed. The rotating brush and a high-speed gas jet work together to deagglomerate the fly ash and disperse it to the same approximate particle size as the original distribution. The disperser, however, is not expected to effectively redisperse submicron particles, which means that the particle size distribution from the disperser will be somewhat biased toward larger particle sizes. Filtration behavior for the redispersed ash should not be significantly biased, though, since submicron particles make up less than 3% of the total mass of typical fly ash.

Initial tests with this system revealed that a significant amount of fly ash was deposited on the walls of the filter holder prior to reaching the fabric. Part of the problem was attributed to particle charging during the dispersion process when dry filtered air was used as the carrier gas. The system was then modified so that the carrier gas was first bubbled through a water bath to provide humidification. This provided about 1.3% humidity to the carrier gas and greatly reduced deposition due to electrostatic effects. Some ash is still deposited on the walls of the filter holder, but the percentage deposited is fairly constant from test to test. The ash feed rate is adjusted to provide a dust concentration of 3 grains/scf reaching the filter.

Sight ports on both the clean and dirty sides of the filter allow visual observation of dust cake buildup during testing. By using various combinations of front and back lighting, small pinholes in the fabric and dust cake are easily detected. The relative thickness of the dust cake can be determined by using intense front lighting and visually monitoring the clean side of the filter. The absence of detectable light on the clean side of the filter indicates that there are no pinholes and that the dust has completely covered the fabric. The fabric sample is mounted in a separate frame and weighed before each test. Following the test, the fabric and frame are removed and weighed with the dust cake intact for determination of the exact areal dust cake weight. Tests were started with new woven glass fabric of the same type as was used in the pilot conditioning tests. However, several tests were conducted using 150 mesh stainless steel screens in place of the fabric. The tests reported here did not include any fabric cleaning. Tests typically were run at an air-to-cloth ratio of 2 ft/min for a duration of 3 hours at ambient temperature. Fly ashes used in the bench-scale tests included conditioned and baseline ashes which were collected from the previous pilot runs.

Respirable mass emissions for South Hallsville and Monticello bench-scale tests are shown in Figures 4 and 5. Superimposed on these figures are the emissions during the first 3 hours of the pilot-scale tests. The emissions curves are remarkably similar considering that the bench-scale tests were at an air-to-cloth ratio of 2 ft/min with redispersed fly ash at ambient temperature compared to a ratio of 3.3 ft/min at 300°F for the pilot tests. The lower air-to-cloth ratio of 2 ft/min was specifically chosen to verify that differences between conditioned and nonconditioned ashes were not simply due to the higher velocity. From Figures 4 and 5, it can be concluded that ash characteristics changed by the conditioning process are retained even when the ash is cooled, redispersed, and filtered. This implies that studying the penetration behavior of redispersed nonconditioned and conditioned ashes under ambient conditions is a valid approach to understanding the mechanisms of improvement.

MECHANISMS OF EMISSIONS REDUCTION

Filter Structure

Particle collection efficiency of a filter is a function of the properties of the particles, the carrying gas, and the filter media. Particle properties which may be important include size, density, shape, electrical charge, and chemical composition. Gas properties which may affect the filtration process include velocity, pressure, temperature, and viscosity. The important filter properties are thickness, fiber diameter, and porosity. Penetration decreases exponentially with increasing filter thickness, decreases with decreasing fiber diameter, and increases with increasing porosity (16). Therefore, it would appear that a good filter should be of adequate thickness, be made up of very fine diameter fibers, and be of low porosity. Pressure drop, however, increases rapidly with decreasing porosity, which is why fibrous filters usually have a porosity of at least 90% (16). Effective filtration results from the use of microfine fibers such as the glass fiber filters which are commonly used for stack or ambient aerosol sampling. These filters are constructed of glass fibers which are about 0.5 to 2 μm in diameter, and have a collection efficiency of 99.99+% for 0.3 μm particles (17).

Differences in the design of fabric filters account for the higher penetration compared to an aerosol sampling filter. Glass fabric filters employ much larger fibers (about 10 μm) than sampling filters, and may be of woven construction which results in large pore openings at the yarn junctions. The differences in structure between a clean dust loading filter and a woven glass fabric are demonstrated with SEM micrographs in Figure 6. While the SEM micrographs clearly show the fabric filter structure, they do not reveal straight-through pores. Typical pore size for the woven glass fabric, however, is revealed with optical microscopy as shown in Figure 7. Backlighting reveals straight-through penetration of the light at some of the yarn junctions. Comparison with the 150 mesh overlay, which has an opening size of 104 μm, shows the pore size to be on the order of 100 to 200 μm. To achieve a high collection efficiency with a woven glass fabric, these large pores must be effectively bridged over. Another structural feature of woven glass fabrics that may affect particulate emissions is the tightly packed fibers in the warp yarns.

Tight packing makes these fiber bundles relatively impermeable which means that the gas flow will go around the warp yarns rather than through them as with the texturized fill yarns. Therefore, the actual local gas velocity may be much higher at the yarn junctions of the fabric/dust interface than in the bulk dust cake. Visual monitoring of dust buildup on new woven glass fabric reveals that initially dust is collected around the yarn junctions followed by a covering of the exposed texturized yarns. However, even after 60 minutes of dust buildup, there are still some exposed warp yarns indicating that most of the gas bypasses these yarns. The effect that this may have on penetration behavior will be discussed later.

Penetration Mechanisms

There are two general mechanisms of ash penetration through a fabric filter. The first is direct, or straight-through penetration, in which suspended particles remain in the gas streamline and pass directly through the filter along with the gas. Direct penetration implies that there must be large enough pores in the dust cake/fabric such that particles can avoid collection. The second type of penetration is indirect, or delayed penetration, in which particles were initially collected, but are somehow reentrained and end up in the gas exiting the clean side of the fabric. Indirect penetration can occur by several mechanisms. One possibility occurs when large pores are initially bridged over, but with increased pressure drop and dust cake buildup, the whole pore bridge breaks loose (18). Indirect penetration may also occur if particles are not held rigidly in the dust cake, gradually seeping through the fabric (18). A somewhat similar mode occurs when particles are reentrained from the rear face of the dust cake, which Dennis (19) refers to as rear face slough-off.

Of the penetration mechanisms, it would appear that direct penetration would be the most straightforward to model. Assuming the number and size of pinholes were known, it would be possible to determine the collection efficiency. Although attempts have been made to include pinhole penetration in efficiency prediction modeling (19,20), variables such as seepage, pinhole size, and the number of pinholes make accurate prediction difficult. Prediction of penetration by indirect seepage or slough-off is even more difficult because the extent of this type of penetration is as much a property of the dust as it is the fabric. Cooper and Hampl (21) recognize the importance of seepage in penetration prediction, but admit that the variables controlling the rate of seepage are unknown. Penetration is additionally complex because several mechanisms may be occurring at the same time. Leith (18), in experiments in which ash was tagged for identification, distinguished between straight-through, seepage, and pinhole plug penetration. He determined that straight-through penetration decreased rapidly with dust cake buildup, but seepage and pinhole plug penetration increased.

The bench-scale tests also revealed at least two separate penetration mechanisms. Fine particulate emissions were high during an initial filtration period for all tests regardless of the ash or conditioning. This represents the bridging process of the larger pores. During this process, visual monitoring of the clean side of the fabric revealed straight-through light transmission. For the ashes with low penetration, emissions continued dropping for approximately 40 minutes, after which they remained at a constant low level. At this point, light transmission through the filter was still obvious, but it was diffused rather than direct transmission. For the ashes with high penetration, emissions leveled off after only 10 to 20 minutes and, in some cases, actually increased with increased filtration time. For some tests, such as with the nonconditioned Monticello ash, numerous pinholes developed in the dust cake which were not readily bridged over with increased dust cake buildup, as shown in Figure 8. Pinholes are easily detected on both sides of the fabric. On the dirty side, a small mound builds around the hole as some dust is deposited when the gas is accelerated through the pinhole. Monitoring the clean side of the fabric with backlighting clearly shows direct light transmission through the pinholes. When pinholes are present, it follows that direct penetration of the dust contributes to high emissions. However, indirect seepage may be occurring simultaneously. If the amount of the gas flow through pinholes can be determined, assuming a low collection efficiency for the pinhole (e.g. 50%), the contribution to emissions from pinhole penetration can be estimated.

Pinhole Penetration

The laminar flow pressure drop across a capillary tube is given as:

$$\Delta P = 32 \mu L V_c / D_c^2$$

where μ = gas viscosity, L = tube length, V_c = average capillary velocity, and D_c = tube diameter. L is replaced by:

$$L = W / \rho_p (1 - \epsilon)$$

where W is the dust cake weight per unit area, ρ_p is the particle weight density, and ϵ is the dust cake porosity. K_2, the specific dust cake resistance (22), is defined as:

$$K_2 = \Delta P / W V_o \quad \text{or} \quad \Delta P = K_2 W V_o$$

where V_o is the filter face velocity or air-to-cloth ratio. Substituting for ΔP leads to the dimensionless relationship:

$$V_c / V_o = 311.6 K_2 (1 - \epsilon) \rho_p D_c^2 / 32 \mu$$

where units are given as:

K_2 = inches of H_2O-ft-min / lb
D_c = ft
ρ_p = lb / ft^3
μ = lb-sec / ft^2
311.6 = coefficient to reduce K_2 to units of sec / ft

Assuming that flow through a pinhole can be modeled by laminar capillary flow, this equation gives the ratio of pinhole velocity to face velocity as a function of K_2 and the pinhole size, D_c, as shown in Figure 9. Note that the velocity ratio V_c/V_o does not depend on the dust cake weight, W, but is directly proportional to K_2. Typical pinhole diameters observed in the bench-scale tests ranged from 100 to 400 μm. From Figure 9, pinhole velocity can be over 1000 times greater than the face velocity, which is similar to results reported by Dennis (19). High pinhole velocity leads to higher emissions because of the greater gas flow. It would also appear that higher velocity could lead to scouring of pinhole sides making the pinhole even larger and further increasing emissions. A lower pinhole velocity will result in reduced emissions and should facilitate the bridging over of the pinhole. Therefore, if pinhole velocity can be decreased by reducing K_2, lower emissions are expected.

The fraction of the gas flow passing through pinholes can be quantified by solving the capillary flow equation for flow rate and making substitutions as before, which leads to a second dimensionless relationship:

$$Q_c\ N\ /\ V_o = 311.6\ \pi\ K_2\ \rho_p\ (1 - \epsilon)\ D_c^4\ N\ /\ 128\ \mu$$

where:

Q_c = pinhole capillary flow rate, ft^3 / min
N = number of pinholes / ft^2 of fabric

The term Q_c N / V_o is the dimensionless fraction of the flow passing through pinholes and is represented graphically in Figure 10. The figure shows that a single 400 μm pinhole per ft^2 of fabric can account for 5% of the flow. Assuming 50% collection efficiency for the pinhole, the fabric filter collection efficiency would be reduced to 97.5%. Because of the fourth power dependence on pinhole diameter, penetration through a 100 μm pinhole would be 256 times less than the penetration through a 400 μm pinhole, and the collection efficiency would be 99.99%. Large pinholes are highly undesirable and any steps that can be taken, such as making K_2 smaller or changing other dust characteristics, to reduce pinhole size or prevent pinhole formation will lead to lower particulate emissions.

Reentrainment Penetration

Some of the bench-scale tests have resulted in fairly high emissions without any trace of pinholes. After about 1.5 hours of dust cake buildup on the fabric, normally no light can be detected penetrating the filter unless pinholes are present. Even with very intense lighting of the dust cake, the clean side of the filter appears completely dark, indicating the absence of any large pores in the dust cake. In spite of the dust cake completely covering the fabric and no evidence of pinholes, emissions can be quite high. Penetration under these conditions must be attributed to reentrainment of dust at the fabric/dust cake interface. It would appear that the opportunity for this reentrainment would be greatest at points where the gas velocity increases to bypass the impermeable fill yarns. At these points, the pore openings may by 10 or more times the median particle size which means there is a dust cake surface without any support, except for that provided by particle cohesion. Therefore, particles can be reentrained as the gas exits the dust cake. If there were no particle cohesion, 10 μm particles would easily be reentrained since the terminal gravitational settling velocity in ambient air is only 0.6 cm/sec compared to the average gas velocity through the dust cake of 1.7 cm/sec for a 2 ft/min face velocity and a porosity of 60%. Study of entrainment behavior in fluidized beds has shown that entrainment is not significant until the gas velocity is considerably higher than the terminal particle velocity; however, once entrainment begins, it increases roughly by a factor of 100 for doubling of gas velocity (23). Applying this concept to the fabric/dust cake interface, if the local gas velocity increases by only 50% to bypass the fill yarns, the potential for seepage or reentrainment is clear.

Critical Ash Characteristics

Mechanisms by which particulate emissions might be reduced include bridging over of large pores in the fabric and dust cake, and reducing or preventing reentrainment. A thin but continuous unflawed dust cake should provide virtually 100% collection efficiency (19,20). Some of the bench-scale tests also show that nearly 100% collection efficiency can be achieved with an average dust cake thickness of only 500 μm. Therefore, the collection efficiency of the continuous dust cake is of little concern. Rather, direct penetration through pinholes and reentrainment must be controlled to reduce particulate emissions. The general property of the dust that describes its ability to bridge large pores and resist reentrainment is the bulk cohesion of the fly ash. This is a term which we have previously referred to as cohesiveness (12,13) and others have called cohesivity (24). In either case, this is not an exact parameter which can be easily measured.

There are many parameters which describe some aspect of bulk cohesion such as the angle of internal friction, angle of repose, angle of wall friction, angle of rupture, and angle of slide (25). Some of these terms describe geometric characteristics of powders or granular material in bin flow. A number of tests are available to measure the shear or tensile strength of bulk powders (26,27), although many of these tests were developed for soil mechanics studies and may not accurately describe the behavior of fine powders such as fly ash. We have previously reported a rough method to determine the relative cohesive strength of fly ash (12). A fly ash sample is compacted into a pellet in a high-pressure press and the breaking strength of the pellet is a measure of the bulk cohesion. Results have shown that ashes for which emissions are greatly reduced by conditioning also show significantly increased cohesive strength with conditioning. While the method did show that conditioning increases cohesive strength, there are two disadvantages. First, some ashes are so low in cohesiveness that the pellet falls apart, making measurement impossible. Second, since the ash is compressed, the porosity of the test sample is about 30% compared to the dust cake porosity of 60% to 70%. The cohesive strength of the compressed sample may not accurately reflect the cohesive strength of the dust cake. Recently, work has been conducted at Southern Research Institute (SoRI) using a ring shear device to measure the cohesive strength of fly ash (24). SoRI terms the parameter which describes the shear behavior the effective angle of internal friction and reports a range for fly ash from 31° for low cohesive ashes to 46° for high cohesive strength ashes. Several of our ash samples were tested at SoRI by this method. Two ashes which have proven to be difficult to collect were each determined to have an effective angle of internal friction of 35°. The angle increased to 39.5° for one of these ashes after being conditioned. The other conditioned ash was not tested. Another ash which has proven to be easy to collect had an effective angle of internal friction of 43°. The limited data indicate that ashes which have an effective angle of internal friction of 35° or less are candidates for emissions reduction by conditioning, and if an angle of about 40° or greater can be achieved through conditioning, significant reductions in emissions can be expected. These results, however, should be interpreted with caution because other factors such as particle size may be important. Ashes with a much smaller or larger particle size distribution than the approximate 10 μm mass median diameter for these ashes may not show the same changes in bulk cohesion and emissions with conditioning.

It follows that increasing the cohesive strength of fly ash through conditioning will facilitate the bridging of large pores and impede the reentrainment of ash from the dust cake/fabric interface. When agglomerated ash is reentrained (such as pinhole plugs breaking loose), the more cohesive ashes would tend to remain as agglomerates rather than break up as individual particles. There would be no difference in total mass emissions, but from a health and visibility perspective emissions of larger agglomerates are less of a problem. The increase in pore bridging ability with conditioning was demonstrated in several tests in which 150 mesh screen was used in place of the fabric. The conditioned South Hallsville ash effectively bridged the 104 μm openings to produce low emissions, while relatively high emissions for the nonconditioned South Hallsville ash indicate difficulty in bridging over large pores.

A question exists as to whether or not other ash characteristics such as particle size or electrical characteristics may be changed with conditioning and play a role in the mechanisms of emissions reduction. Previously reported results (12-15) from the pilot-scale tests show that conditioning produced little or no change in the particle size distribution of the aerosol upstream of the baghouse. Measured changes in electrical resistivity show that conditioning lowers the resistivity at 300°F by 0.5 to 1.5 orders of magnitude; however, the change in resistivity is not expected to affect the filtration process unless charge dissipation is a problem. Several tests were conducted to measure the charge level of the fly ash from the output of the aerosol generator with a TSI Model 3068 aerosol electrometer. Although readings were not steady, the indicated charge/mass ratio was on the order of negative 2 X 10^{-8} coulombs/gram, which would be considered a very low net particle charge. Other indications that electrical effects are not significant to the mechanism are that similar percentages of the ash reach the filter for both conditioned and nonconditioned ashes. If either ash were more significantly charged, more ash would tend to collect on the metal walls of the filter holder. Therefore, electrical characteristics and particle size do not appear to have a major role in the reduction of emissions with conditioning. Rather, modification of bulk or particle cohesion best explains the reduction in emissions.

EFFECT OF CONDITIONING ON PRESSURE DROP

The pressure loss across a fabric filter is the result of viscous flow through a porous medium. The magnitude of the pressure drop is dependent on gas properties such as viscosity, density, and velocity, and on filter media properties such as filter thickness, porosity and pore size. Assuming incompressible, steady viscous flow, the pressure drop increases linearly with dust cake thickness and gas velocity:

$$\Delta P = K_2 \, W \, V$$

which is derived directly from Darcy's law. Prediction of K_2 in terms of measurable dust parameters has been attempted, but accurate prediction has not been attained. The Carman-Kozeny relationship is derived from a theoretical capillary model and, assuming spherical particles, takes the form (28):

$$K_2 = 36 \, k \, \mu \, (1- \epsilon) \, / \, \rho_p \, \epsilon^3 \, D^2$$

where k is an empirical constant approximately equal to 5, but the porosity dependence is strictly theoretical. This relationship has been criticized as not being valid for porosities greater than 70%, so alternative models such as the Rudnick and First (29) model have been proposed:

$$K_2 = (18 \, \mu / \rho_p \, D^2) \, \{[3 + 2(1 - \epsilon)^{5/3}] \, / \, [3 - 4.5(1 - \epsilon)^{1/3} + 4.5(1 - \epsilon)^{5/3} - 3(1 - \epsilon)^2]\}$$

Vann Bush at SoRI (30) has recently reported an empirical relationship relating K_2 to porosity for coal fly ash:

$$K_2 = (4 \, \mu \, / \, D^2 \, \rho_p) \, [(1 - \epsilon)/\epsilon] \, [7.5 + 9.1(1 - \epsilon) - 35.8(1 - \epsilon)^2 + 560(1 - \epsilon)^3]$$

where D is referred to as the drag equivalent diameter. In the Carman-Kozeny and Rudnick-First equations, D refers to the actual physical diameter for monosized spheres, but for a particle size distribution such as fly ash, the mass median diameter generally cannot be used for D. A weakness, therefore, in all of these relationships is that the particle diameter term is not a directly measurable parameter. The value of the characteristic diameter in these equations is a complex dependence on the particle size distribution, specific surface area, and particle shape. The strong dependence of K_2 on particle size and porosity, however, is demonstrated by these relationships.

The pilot-scale conditioning tests showed that baghouse pressure drop was significantly reduced compared to baseline tests without conditioning. The bench-scale tests also confirm this reduction in pressure drop. For the South Hallsville ash, K_2 was reduced from 7.8 (inches H_2O-ft-min / lb) for the baseline test to 4.8

for the conditioned ash. The Monticello ash K_2 dropped from 8.3 for the baseline to 6.6 for the conditioned ash. Since the particle size distribution is not significantly changed with conditioning, a drop in K_2 should correspond to an increase in porosity. Porosity measurements for both the conditioned and nonconditioned ashes indicate that this is the case. Since the dust cake thickness following the bench-scale tests was only about 1 mm, direct measurement of the porosity was not possible. An alternative method involved carefully transferring the cake ash into a graduated cylinder for determination of the volume of a given mass of ash. Along with the particle density, measured with a helium pycnometer, this provided the data to calculate porosity. Porosities measured by this method were similar to the porosity determined from measuring the volume and mass of a bulk ash sample under minimum compaction. Porosities for the conditioned and nonconditioned South Hallsville ashes were 71% and 64%, respectively. Monticello ashes had measured porosities of 68% and 58% for the respective conditioned and baseline samples. The K_2 and porosity values for these ashes are shown in Figures 11 and 12 along with the Carman-Kozeny, Rudnick-First, and SoRI models. The equations were solved for the characteristic particle diameter at the measured K_2 and porosity values for the nonconditioned ashes, and therefore converge at this point. Note that the measured decrease in K_2 with porosity for the South Hallsville ash is close to the value predicted by the Carman-Kozeny curve. The measured drop in K_2 for the Monticello ash, however, was much less than that predicted by the models, as shown in Figure 12. An explanation is that the actual dust cake porosity may have been greater than the measured porosity for the nonconditioned sample due to packing during handling. The relative susceptibility of the ashes to packing is indicated by the porosity after settling in a graduated cylinder which is tapped on the side until the ash settles to a final packing density. Dust porosities after settling were 43% and 34% for the respective conditioned and baseline South Hallsville ash, but only 40% and 29% for the respective conditioned and baseline Monticello ash. The very low value of 29% indicates this ash is highly susceptible to packing and suggests that conditioning may reduce pressure drop by inhibiting packing of the residual dust cake due to vibration and shock after bag cleaning.

The factors which determine the porosity of a dust cake are not well known, other than the general increase in porosity with decreasing particle size for a variety of bulk powders (22). Particles larger than 30 µm tend to have porosities in the 40% to 50% range, while smaller sized particles tend to form much more porous structures, presumably due to the relationship between cohesive forces and particle mass (22). Very fine particles of about 1 µm or less will typically have porosities of 90% or greater. Pohl (24) recently reported a correlation between dust cake porosity and the bulk cohesion of the dust, the more cohesive ashes forming the more porous dust cakes. The Carman-Kozeny, Rudnick-First, and SoRI models predict a reduction in K_2 by a factor of 2 to 3 by increasing the porosity from 60 to 70%. Therefore, if the dust cake porosity can be increased following an increase in the bulk cohesion through conditioning, a reduction in pressure drop is expected.

SUMMARY

Pilot and bench-scale results show that fine particulate emissions from a woven glass fabric filter can be greatly reduced by conditioning with small amounts of SO_3 and ammonia. The greatest reduction in emissions is expected for ashes which are somewhat difficult to collect because of low bulk cohesive strength. Conditioning reduces emissions by enhancing the ability of the ash to bridge large pores and pinholes, by inhibiting the reentrainment of ash from the dust cake/fabric interface, and by reducing K_2 which results in lower pinhole velocity. These three mechanisms of emissions reduction are a direct result of increasing the cohesive strength of the ash due to conditioning. An increase in the cohesive strength also leads to increased dust cake porosity. Pressure drop models predict that a relatively small increase in porosity will significantly reduce K_2. Measured increases in porosity with reduced K_2 values for conditioned ashes confirm that ammonia and SO_3 conditioning reduces pressure drop.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

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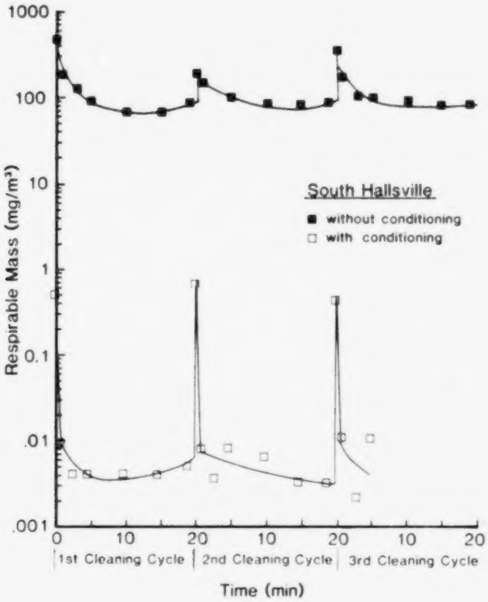

Figure 1. Respirable mass emissions as a function of time for South Hallsville Texas lignite pilot-scale tests with and without ammonia and SO_3 conditioning.

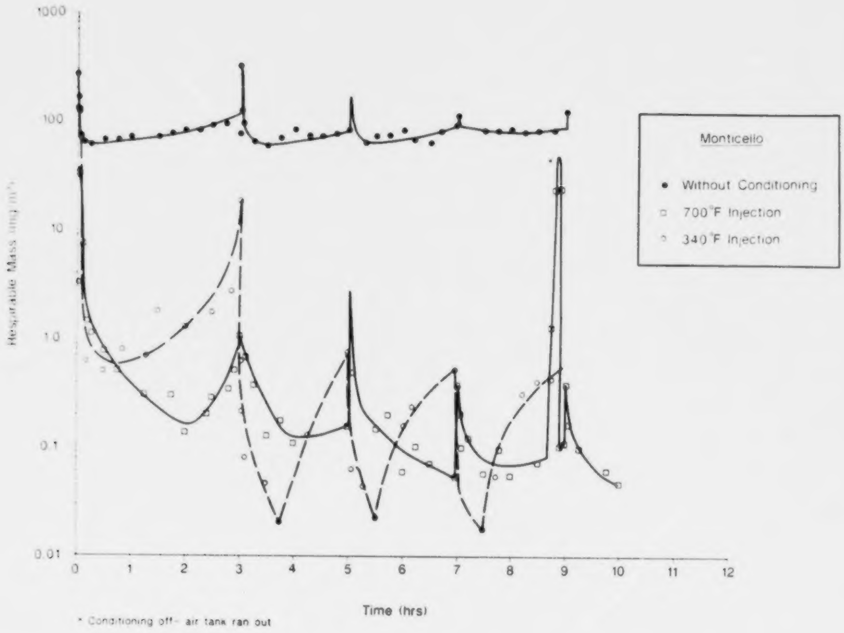

Figure 2. Respirable mass emissions as a function of time for Monticello Texas lignite pilot-scale tests with and without ammonia and SO_3 conditioning (bags were cleaned at 3, 5, 7, and 9 hrs).

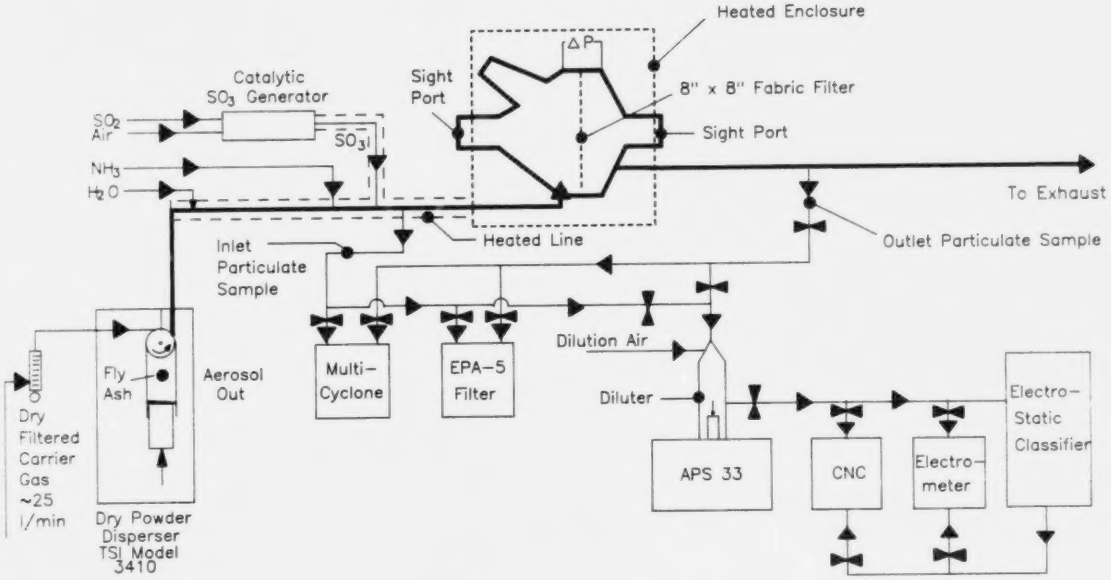

Figure 3. Schematic of bench-scale fabric filtration system.

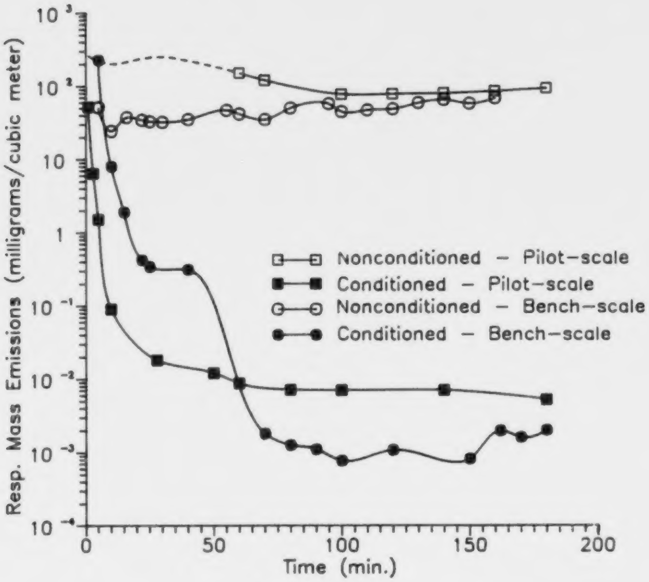

Figure 4. Respirable mass emissions as a function of time, comparing bench-scale and pilot-scale conditioning tests for South Hallsville Texas lignite.

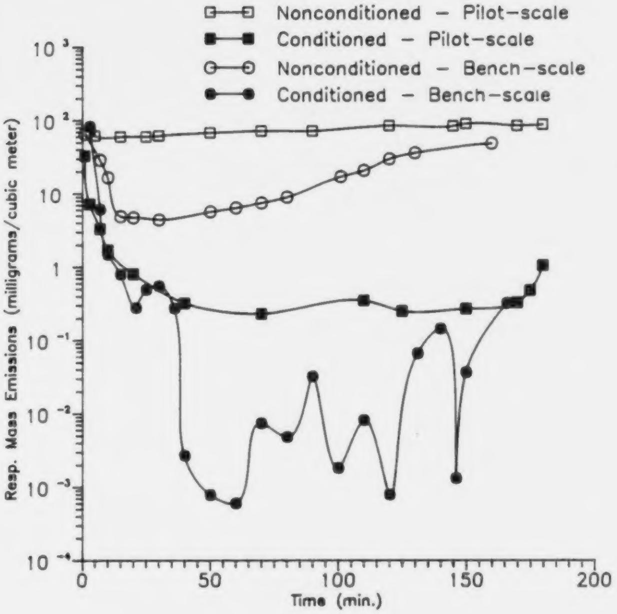

Figure 5. Respirable mass emissions as a function of time, comparing bench-scale and pilot-scale conditioning tests for Monticello Texas lignite.

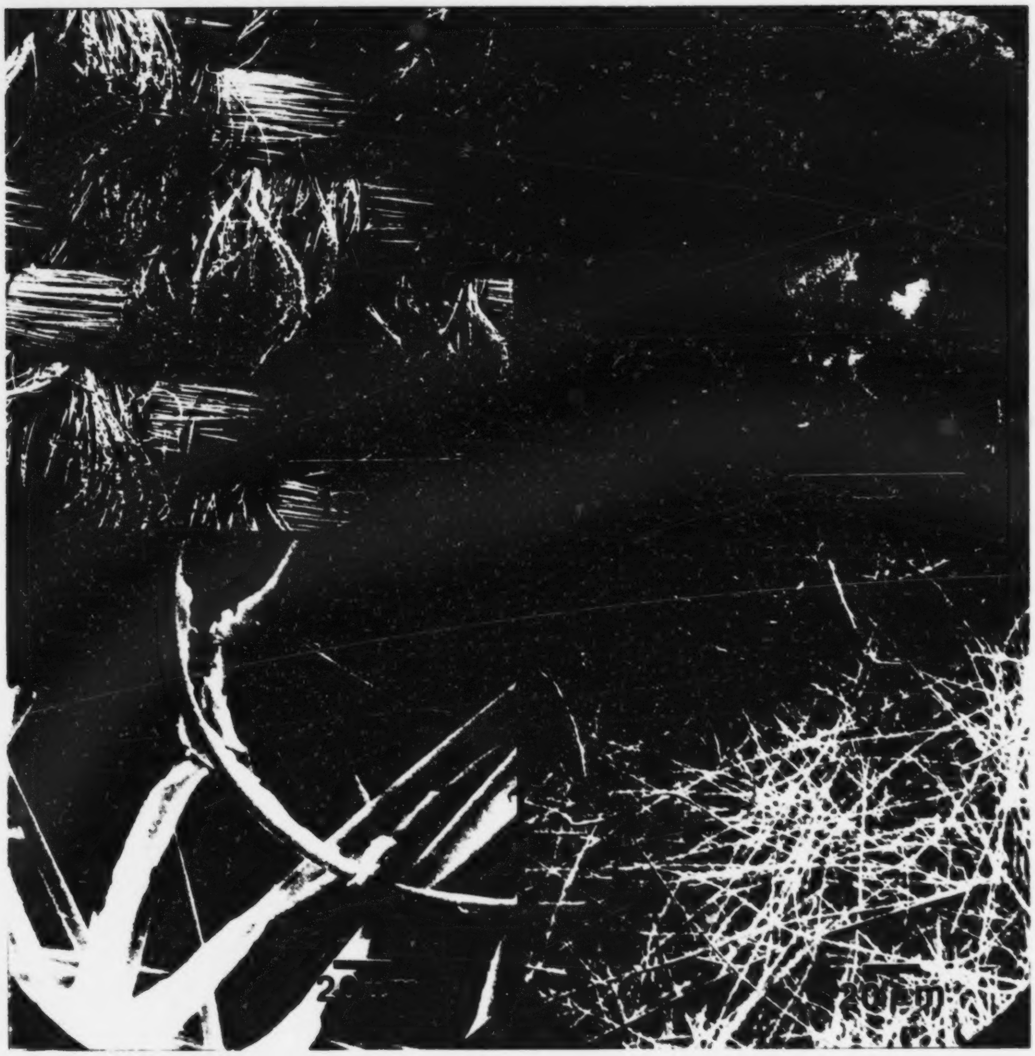

Figure 6. Structural comparison between a 601E fiberglass fabric and a 934AH Whatman glass filter (magnification: 30x and 500 x).

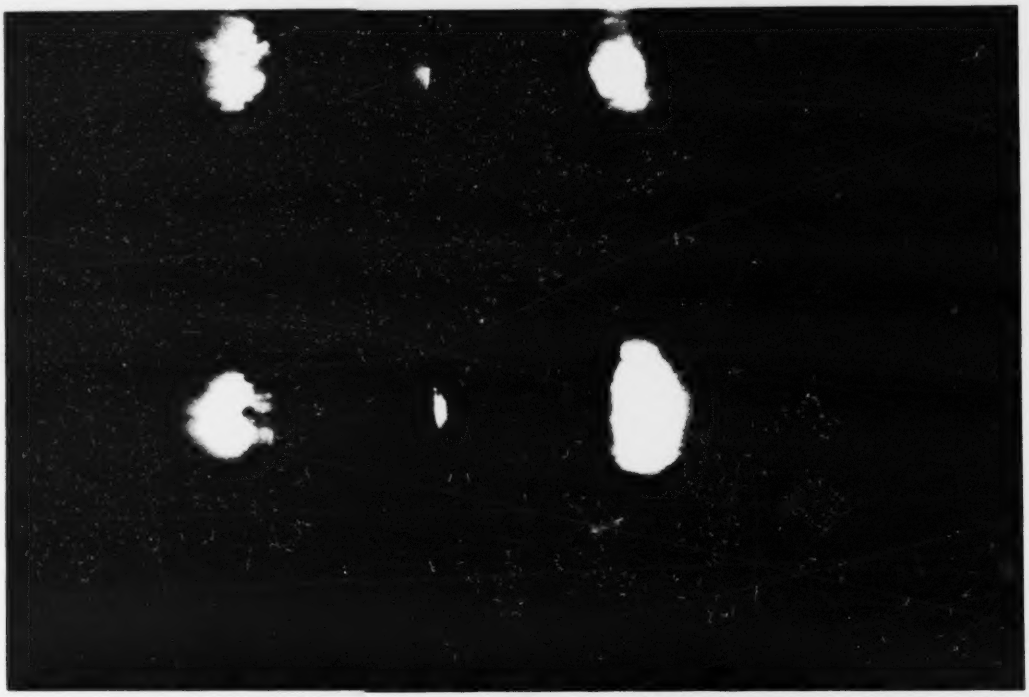

Figure 7. Comparison of pore size between a 601E fiberglass fabric with a 150 mesh stainless steel screen.

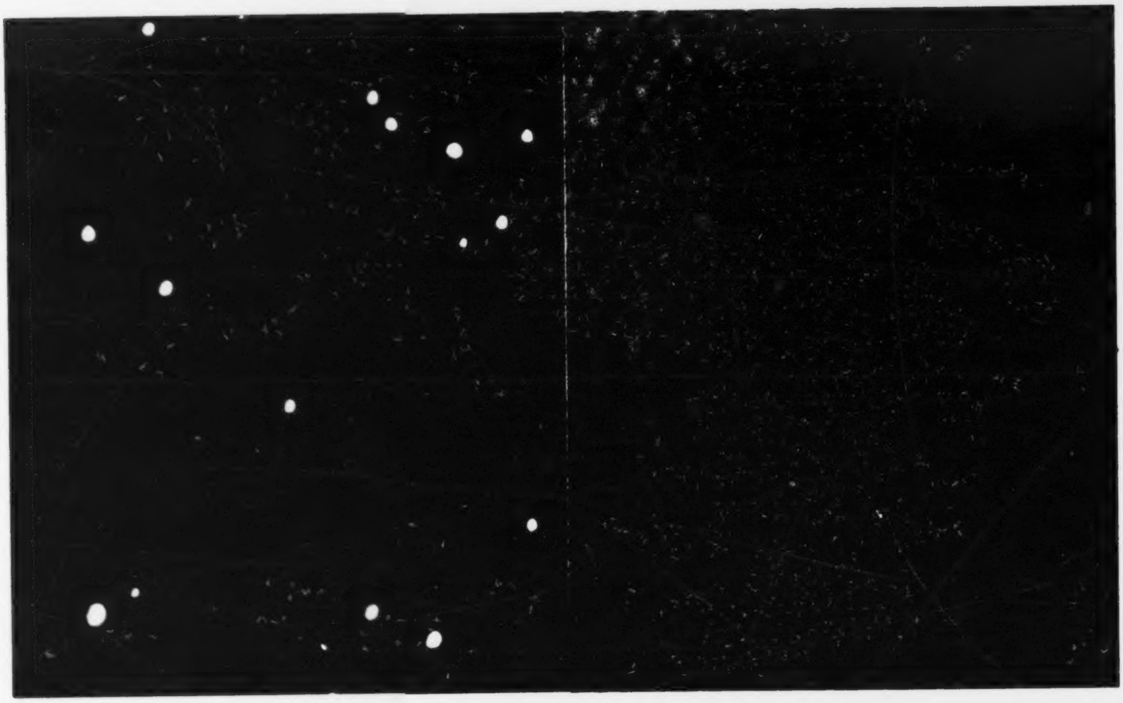

Figure 8. Dust cake pinholes, with and without backlighting, at the end of a bench-scale test using nonconditioned Monticello Texas lignite.

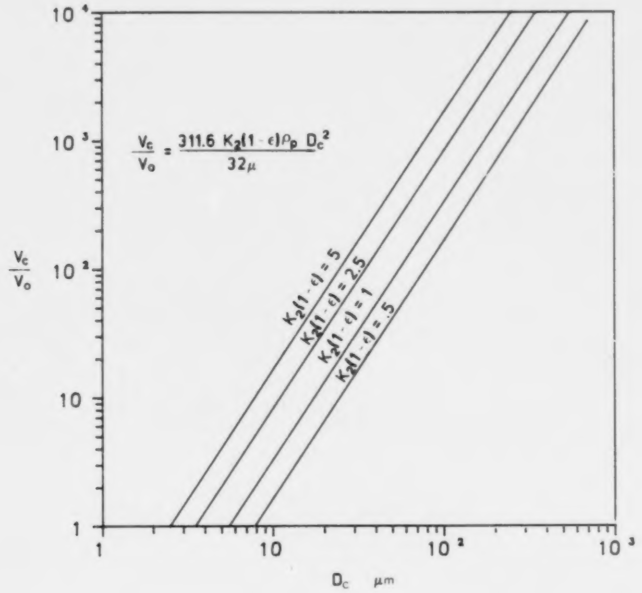

Figure 9. Dimensionless pinhole velocity as a function of pinhole size.

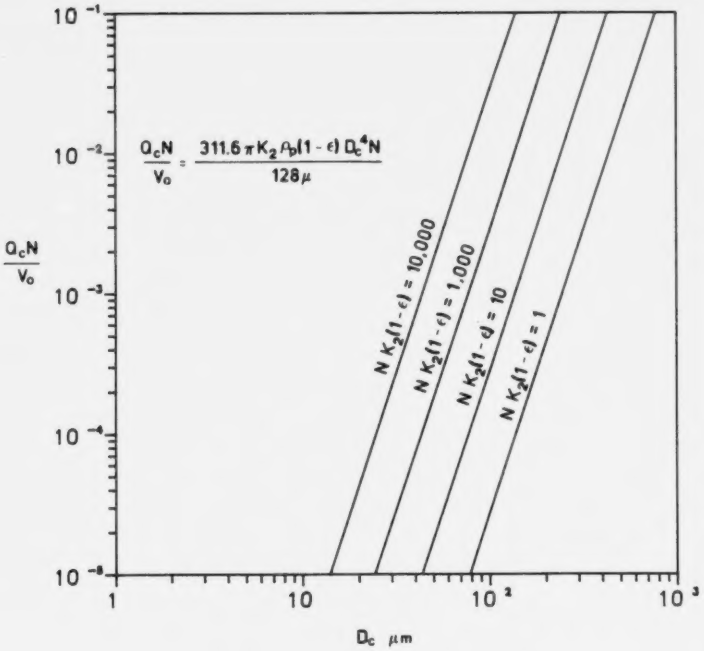

Figure 10. Dimesnionless fraction of gas flow passing through pinholes as a function of pinhole size.

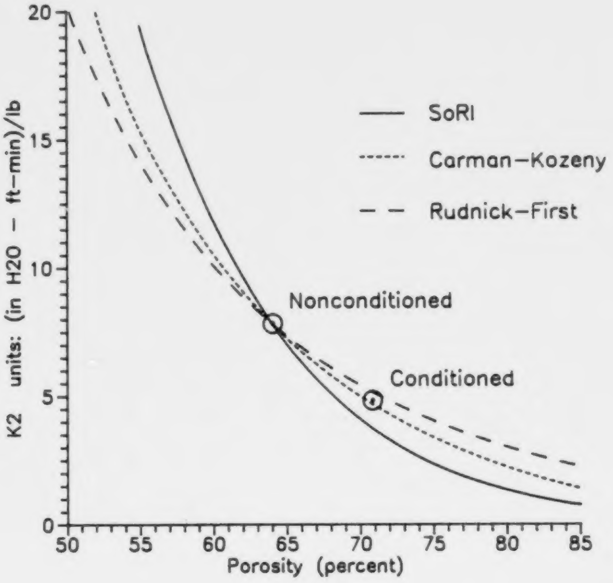

Figure 11. Specific dust cake resistance (K_2) as a function of porosity for three different models using South Hallsville Texas lignite.

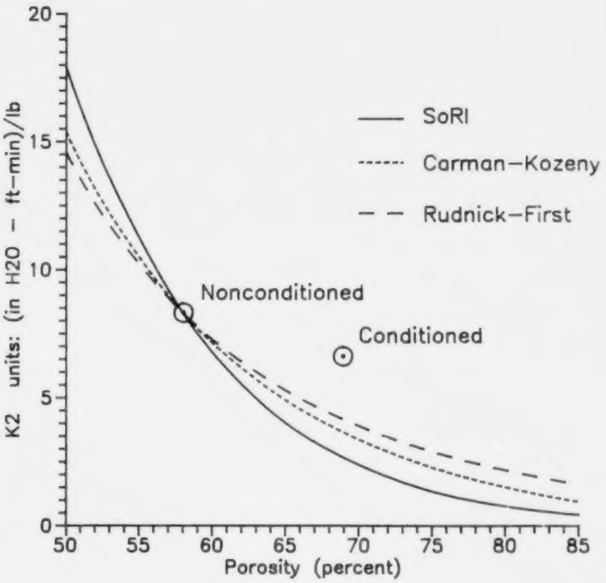

Figure 12. Specific dust cake resistance (K_2) as a function of porosity for three different models using Monticello Texas lignite.

ADVANCES IN ELECTROSTATICALLY STIMULATED FABRIC FILTRATION

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This paper has been reviewed in accordance with the U.S. Environmental Protection Agency's peer and administrative review policies and approved for presentation and publication.

ABSTRACT

The status of advanced electrostatically stimulated fabric filtration (AESFF) is discussed. A mathematical model has been developed that explains the formation of the non-uniform deposit of particle matter that results from the application of AESFF to a fabric filtration bag. The model also explains how the non-uniform deposit causes a reduction of pressure drop as compared to conventional filtration. The model allows the prediction of the relative pressure drop reduction that results. The application of AESFF to fabric filtration results in a significant reduction in particle penetration as compared to conventional filtration. Finally, some of the technical problems that remain to be solved are discussed.

INTRODUCTION AND BACKGROUND

Advanced electrostatically stimulated fabric filtration (AESFF) is a technology developed by the U. S. Environmental Protection Agency that has the potential for providing a significant advance in fabric filtration capability. It will simultaneously allow more economical operation, and decreased particle penetration through the filtration media. This paper summarizes the status of AESFF.

Fabric filtration in the form of baghouses, one of the principal technologies for the removal of particulate matter from gas streams, has been used for many years by various industries. In recent years, in response to increasingly stringent air pollution control requirements, baghouses have gained more prominence, especially in the electric power industry. Electrostatic precipitators are used in the majority of utility plants, however, in the United States in 1984 there were over 110 baghouses associated with more than 20,000 MW of electric generating capacity.[1] In a typical application, gas flow is arranged for inside-to-outside flow with the dust collected on the interior of the bag. The work described in this paper is for inside-to-outside flow which is used for all coal fired utilities in the United States and many of the high volumetric gas flow applications in other industries.

For inside-to-outside-flow fabric filtration bags, woven fabrics are usually used. Woven fabrics have pores that are large in comparison to the size of the particulate matter or dust that is to be filtered. Their filtering action depends on the building up of a dust cake, which actually performs the filtration, on the surface of the fabric.

To make a practical filtration system, some means must be introduced for cleaning the accumulated dust from the bags to maintain the pressure drop within the fan operating limits. For the inside-to-outside flow method this cleaning is done by reversing the gas flow, mechanical shaking, or some combination of both. Filtration is a continuing process consisting of dust accumulation and cleaning, followed by another period of dust accumulation. The bag is not cleaned to bare fabric. It maintains a residual dust cake which tends to cause a pressure drop increase slowly over a long period of time during the life of the bag, which is about 2 to 5 years.[1]

A major consideration in the design and operation of baghouses is the maximum increase in pressure drop, as the dust cake builds up. If this pressure drop could be minimized, there would be a significant fan power savings. Another aspect of this filtration cycle pressure drop is the limit on gas flow. If the pressure drop could be reduced, the allowable gas flow could be increased. Reviewing the data that have been reported on electric power industry baghouses [1] suggests that the typical gas-to-cloth ratio for inside-to-outside flow is about 0.01 m/s. For a full size utility bag 9 to 10 m long [2], with a diameter of 0.3 m, flow through each bag would be limited to about 6 m^3/min. The gas flow from a 500 MW capacity electric generating plant, assuming a reasonable gas flow rate of 100 m^3/min/MW of electric generating capacity, would require about 8000 bags. Reducing the number of bags could be the basis of a worthwhile capital saving.

Considerable work has been done on limiting the pressure drop rise during the filtration cycle. Much of the work has been related to making engineering changes to the fabric, improving the flow distribution and uniformity in the baghouse, and applying electrostatics.

Much research and development has been devoted to the application of electrostatics.[3] It has long been recognized that electrostatic charges and forces play a role in fabric filtration.[4] It has been claimed that the effects of electrostatics, mainly in the triboelectric properties of dusts and fabrics, are very important, if not critical, in controlling filtration properties. The work that was done by the various researchers involved:

- Charging particles prior to collection.
- Collecting particles in an electric field.
- Combining particle charging and collection.

From the reported work have come reports of significant pressure drop reductions.

Recent significant work that has made use of electric fields perpendicular to the flow, called electrostatically stimulated fabric filtration (ESFF), has been developed under the auspices of the U. S. Environmental Protection Agency. The ESFF concept makes use of electrodes which are placed axially, parallel to each other on about 20 mm centers. Electric fields of about 200-800 kV/m between adjacent electrodes are used; operation is below corona onset voltage. The electrodes were either in a harness (wrapped around a bag) or stitched or woven into the bag fabric. The work on ESFF has ranged from small laboratory baghouses, operating at temperatures up to 150°C, to fairly substantial field pilot units.[5] Pressure drop reductions were usually about 40 to 50%.

A system for providing integral particle charging and collection in a combined electric and flow field (as shown in Figure 1) was developed as an outgrowth of earlier work on ESFF. In this system, called AESFF, the corona discharge wire is axially located within the fabric bag, and the electric field is developed between the corona discharge wire and the bag. The bag is made conductive by including conductive yarns in the fabric during the weaving process. Alternatively, if the particulate matter is low in resistivity, it could make the bag sufficiently conductive so that it would be unnecessary to weave conductive yarns into the fabric. The generation of corona for charging and development of the electric field requires only one power supply. Except for its porous walls and closed ends, the device is similar to a wire-pipe ESP.

The system was tried on a small scale in a laboratory single-bag experimental baghouse at the U.S. Environmental Protection Agency's Air and Energy Engineering Research Laboratory at Research Triangle Park, North Carolina. A non-uniform deposition was observed to result in a reduced pressure drop over that of both conventional filtration and ESFF. It was found that pressure drops of 10 to 30% of that of conventional fabric filtration were being achieved.[5]

A decision was made to proceed with a field pilot plant evaluation. The pilot unit contained six bags, each 7.3 m long and 0.2 m in diameter. The spreader-stoker boiler, from which the slipstream was taken, typically had a large amount of unburned carbon in the fly ash. Preliminary measurements indicated that the resistivity was 1 x 10^5 ohm-cm or less, sufficiently low so that three of the six bags, which were woven of a PTFE finished glass fabric, had no provisions to increase their conductivity. The other three bags had stainless steel yarn woven into them longitudinally (axially). After 4 months of tests it was found that it was possible to operate at a pressure drop that was only 20 to 30% of that of conventional filtration. It was also shown that, for low resisitivity dust, there was no difference in performance between the non-conductive bag, which depended on particulate matter conductivity, and the bags that had conductive yarn woven into them.[6,7]

A second pilot evaluation was undertaken at the EPA pilot baghouse at Southwestern Public Service where some of the earlier ESFF field pilot work had been done. The fly ash, from a pulverized-coal-fired boiler, had a considerably higher electrical resistivity than did the spreader-stoker fly ash. The bags that were used had woven-in stainless steel yarns, connected electrically to ground. This pilot unit is currently being operated.[8]

EXPERIMENTAL PROGRAM FOR DEVELOPMENT OF MODEL

A single-bag experimental baghouse at the EPA laboratory facilities at Research Triangle Park, North Carolina, was set up as shown in Figure 2 as a means of gathering the data to be used to develop an engineering design methodology for AESFF. Dirty-particle-bearing gas is introduced through the hopper and flows up into the bag, with the particulate matter deposited on the inner surface. The cleaned gas flows through the fabric and out of the baghouse (inside-out filtration). The fabric used for the filter bags was of woven glass fibers similar to that used in the pilot plant tests. Some tests used the fabric in standard filtration form, with unmodified electrical conductivity. The electrical conductivity of these fabrics depended mainly on that of the materials plus the particulate matter that coats them when used. Other tests used a bag of the same material, except that the conductivity had been modified by having stainless steel yarns woven into the fabric. The yarns, running axially on 20 mm centers, each consists of 90 fibers of 316L stainless steel of 22 μm diameter.

The bags could be opened along the seam to allow removal from the baghouse and placement on a flat surface for examination and testing of the dust cakes that were deposited. The same bags were used over and over again for the various tests. The particulate matter that was used had been obtained, after collection in a baghouse, at the SWPS Harrington Power Station in Amarillo, Texas, which is the location of the second pilot plant. The size distribution of the undispersed SWPS fly ash was determined by outfitting a bag with a calibrated inertial cascade impactor inserted at mid-height. Under isokinetic sampling conditions, the geometric mass mean diameter was 3.3 μm, and the geometric standard deviation was 2.9. The charge-to-mass ratio of the redispersed SWPS fly ash was measured using this same test bag. On a total basis the fly ash was found to be almost electrically neutral.[5]

The baghouse arrangement, shown in Figure 2, was used to determine the effects of varying voltage, current, resistivity, temperature, and gas flow. Voltage was varied by changing the setting of the DC power supply. The voltages applied to the corona discharge wire were negative and ranged from 23 to 50 kV. The current was varied by changing the diameter of the corona discharge wire for different tests--sizes were 2.4, 3.2, 6.4, and 12.7 mm. Resistivity of the SWPS fly ash was changed by varying temperature and moisture. Negative corona was used because of the tendency of positively energized electrostatic precipitators to spark at lower voltages. This is the reason that negative polarity for the corona is used almost exclusively in modern precipitators. The temperatures were 21, 82, and 149°C. A temperature of 149°C is similar to that of many coal-fired steam-electric power plants. Moisture levels were varied from about 1% to slightly more than 9%. Moisture levels of 6 to 9% are the range in which most coal-fired boiler gases operate.[6]

After each test run the bag would be removed and opened, and measurements made of areal density, dust cake thickness, and permeability. Afterwards the bag would be cleaned and put back into the baghouse for the next test. The average clean bag drag was 660 Pa/m/s. Drag (resistance-to-flow) is the reciprocal of permeability.

Figure 3 is a typical set of test results. For this test run the electric field was 7.5×10^5 kV/m, the average current density was 3×10^{-5} A/m^2, and the gas-to-cloth ratio was 0.02 m/s. The axis labeled "Distance along Electrode" is that portion of the fabric filter bag near the corona discharge electrode between the lower oval weight and the upper oval termination of Figure 2. The drag, areal density, and dust cake thickness were measured independently. The dust cake density was computed from the areal density and dust cake thickness.

The preliminary conclusions of the single-bag tests were:

1. The drag decreased with increasing distance along the electrode--the highest drag was at the inlet.
2. The drag always decreased with decreasing areal cake density and dust cake thickness.
3. There was no correlation between drag and dust cake density. Sometimes the dust cake density decreased with decreasing drag. At other times it would remain relatively constant or even increase.

In addition, a very useful conclusion that was drawn was that the face velocity, which is the permeability (or typically inverse of the drag), generally increased linearly as a function of position, as shown in Figure 4.

ENGINEERING DESIGN MODEL DEVELOPMENT

The mathematical model which had been developed [5], for fabric filtration with integral particle charging and collection in a combined electric and flow field, had to consider two types of particle collection--the portion due to gas as porous flow through the fabric and the portion due to electrostatic precipitation. The particle collection due to the electrostatic precipitation effect required knowing if the flow was turbulent or laminar. If turbulent, the particle collection would be expressed by the Deutsch-Anderson equation. If laminar, a simpler collection relationship could be used.[6] The inlet velocity for a full size commercial bag is of the same order as the gas flow velocity in an electrostatic precipitator, which usually has turbulent flow. This suggests that a fabric filtration baghouse might also have turbulent flow, especially near the inlet.

The linearity of face velocity with a position, as shown in Figure 4, allowed a relationship to be developed for the velocity as a function of position in the bag. This subsequently allowed the Reynolds number to be computed as a function of the product of the average gas/cloth ratio, $\overline{V_f}$, and bag length, L, for parameters of position in bag, z, as seen in Figure 5. Fabric filtration bags of the length and average gas/cloth ratios used in commercial practice operate in the region above the turbulent-to-laminar transition, commonly accepted as a Reynolds number of 2100. The small bags used in the laboratory were operating below a Reynolds number of 2100 and were consequently operating in the laminar regime. Therefore for a practical model, the electrostatic precipitation collection was assumed to be taking place in the turbulent flow regime.

The two particle collection mechanisms--porous flow and electrostatic precipitation--require that a bag be divided into a number of length increments. For each increment the amount of particulate matter collected is

$$N_{j-1}\,[(1 - H_j)\,G_j + (1 - G_j)\,H_j] = N_j \tag{1}$$

in which N_{j-1} is the total amount of particulate matter entering the j increment, and N_j is the amount of particulate matter exiting to the j + 1 increment.

In Equation (1), G_j, which is the fraction of the particulate matter collected due to porous flow, is

$$G_j = \frac{(V_{j-1} - V_j)(\pi b^2/4)}{(V_{j-1})(\pi b^2/4)} = \frac{V_{j-1} - V_j}{V_{j-1}} \tag{2}$$

in which V_{j-1} is the velocity of the gas entering the j increment, and V_j is the velocity of the exiting gas. The quantity $\pi b^2/4$ is the cross sectional area of the bag having a diameter b. The velocity of the gas entering and exiting the increment is computed by the expression for velocity, previously mentioned, that had been based upon the assumption of a linear increase in face velocity as a function of position.

The fraction of the dust collected by electrostatic precipitation, H_j, makes use of the Deutsch-Anderson collection equation

$$H_j = \sum_{i=1}^{m} \{1 - \exp\,(-\overline{SCA_j}\,w_{ji})\} \tag{3}$$

in which the average specific collector area, $\overline{SCA_j}$, is the ratio of the collecting area to the gas flow, and w_{ji} is the electrical migration velocity. The index i is the ith particle size increment. The derivation of the Deutsch-Anderson equation assumes that the migration velocity is a constant for all particles, which requires that the size distribution be divided into a number of individual size increments.

In Equation (3), the average specific collector area, $\overline{SCA_j}$, is

$$SCA_j = \frac{\pi b L/l}{\{(V_{j-1} + V_j)/2\}\pi b^2/4} = \frac{8L}{bl(V_{j-1} + V_j)} \tag{4}$$

in which b is the bag diameter, L is the bag length, l is the number of length increments, and the inlet and outlet gas velocities are V_{j-1} and V_j, respectively.

The electrical migration velocity, w_{ji}, is

$$w_{ji} = \frac{2a_i b \epsilon E_1 E_2 J \ln\{(L + z_j)/(L - z_j)\}}{g[bJ \ln\{(L + z_j)/(L - z_j)\} + 32\epsilon E_3 \overline{V_f}]} \times \left\{1 - \exp\left[\frac{-9/16gb \ln\{(L + z_j)/(L - z_j)\}}{a_i^2 r \overline{V_f}}\right]\right\} \qquad (5)$$

in which a_i is the ith increment particle diameter, b is the bag diameter, ϵ is the permittivity of free space, E_1 is the charging electric field, E_2 is the collecting electric field, E_3 is the electric field near the bag surface, J is the average current density, L is the bag length, g is the gas viscosity, r is the particle density, $\overline{V_f}$ is the average gas-to-cloth ratio, and z_j is the location of the mean distance of increment j from the bag inlet.

In the development of Equation (5), it is assumed that the particle enters the bag with zero charge. The charge builds up and approaches saturation as the particle travels up the bag. The time term that appears in the basic expression for migration velocity [6] had been related in the derivation to the expression for gas velocity that had been described previously. Finally, in the development of Equation (5) it is assumed that the only charge carriers are ions.

Figure 6 shows the computation for the non-uniform particulate matter deposition for the bag geometry, electrical conditions, and particle size distribution for the test conditions of Figure 5.

Figures 7 and 8 show the model computation for the two pilot plant tests. The bag geometry and electrical conditions are stated for each. The fly ash, for the field pilot test of Figure 7, is the same as was used for the single-bag laboratory tests: the geometric mass mean diameter was 3.3 μm, and the geometric standard deviation was 2.8. The pilot test, for which the computation of Figure 8 was performed, was for a spreader-stoker furnace whose fly ash had a geometric mass mean diameter of 18.4 μm, and a geometric standard deviation of 5.2.

Of specific interest is the dominant effect of electrostatic precipitation collection within the bag. In conventional (non-electrified) filtration, the collection in each increment would be uniform and would all be due to porous flow.

The mathematical model that has been developed provides a description of the non-uniform deposition that occurs within a bag as a function of bag geometry, particle characteristics, and electrical conditions. In preparing a design, the bag geometry can be established by choice, the particle characteristics are set by the coal and combustion or process conditions, and the remaining are the electrical conditions. The best approach at this time is to make use of the relationship for relating voltage to current density in a wire-pipe ESP. [7]

$$V = V_o + aE_c \left\{ [1 + b^3 J/(\epsilon \mu a^2 E_c^2)]^{0.5} - 1 - \ln \left[\frac{1 + [1 + b^3 J/(\epsilon \mu a^2 E_c^2)]^{0.5}}{2} \right] \right\} \qquad (6)$$

in which V_o is the corona onset voltage, E_c is the corona initiation field at the wire, a is the corona wire diameter, and μ is the ion mobility. The quantities b,J, and ϵ have been previously used in Equations (2), (4), and (5).

By use of Equation (6), the voltage as a function of current density for a bag of a given diameter can be determined. There may be some inaccuracy introduced from the charge that will build up on the surface of the dust layer [5]; however, the inaccuracy that would be introduced is not considered to be excessive. It should be noted from Equation (6) that, for a bag of a given diameter, the current density can be changed by varying the corona wire diameter. The voltage, V, of Equation (6) is converted to electric field by dividing by the wire-to-bag distance. The electric field is then substituted for E_1, E_2, and E_3 in Equation (5).

The model was extended to predict the relative pressure drop of an electrified fabric filtration bag in respect to a bag with a non-uniform deposit. The relationship for drag or resistance to flow, S, is

$$S = k_2 W = \Delta P/V_f \qquad (7)$$

in which k_2 is the specific cake resistance, W is the areal density, Δp is the pressure drop across the dust cake, and V_f is the face velocity. If it is assumed that k_2 does not vary with either dust cake thickness for electrified vs. non-electrified condition, then the average drag for the non-uniform deposit, S_n is

$$S_n = [\sum_{j=1}^{l} \{(1/l)/S_j\}]^{-1} = l[\sum_{j=1}^{l} \{1/(k_2 W_j)\}]^{-1} = l k_2 [\sum_{j=1}^{l} \{1/W_j\}]^{-1} \qquad (8)$$

In Equation (8), l is the number of increments, S_j is the drag for the j increment, k_2 is the specific cake resistance, and B_j is the areal density of the increment.

It follows from Equations (7) and (8) that

$$\Delta P_n/\Delta P_u = S_n/S_u = l[\sum_{j=1}^{l}(1/B_j)]^{-1}/B_u \tag{9}$$

which, for non-uniform and uniform deposition, relates the relative pressure drop, $\Delta P_n/\Delta P_u$, to the ratio of the drags, S_n/S_u. The subscripts n and u denote nonuniform and uniform deposition, respectively.

The relative pressure drop computed by Equation (9) for SWPS, the pilot unit of Figure 7, was 0.44, which compares to a measured value of 0.43.[8] For the spreader-stoker pilot unit of Figure 8, the computed relative pressure drop was 0.22 which compares quite well to a measured value of 0.20 to 0.30.[9]

PARTICLE PENETRATION

Figures 9 and 10 show the particle count for a bag with AESFF on and off, for two different particle size ranges, 0.3 to 1.0 μm and 1.0 to 3.0 μm as a function of time as the dust cake builds up upon the fabric bag surface. The particle penetration is given as counts per minute, as given by a particle counter. Therefore, the results should be used as an indication of relative change in particle penetration. For the 0.3 to 1.0 μm range, the relative reduction in particle penetration was about a factor of 5. For the 1.0 to 3.0 μm range, the reduction in particle penetration was almost 2 orders of magnitude. For both of these tests, the current density was very low, less than 1×10^{-5} A/m2. This is considerably lower than a normal current density, which would usually be at least a factor of 5 to 10 higher. It is expected that, at higher current densities, the relative particle penetration would decrease still further.

REMAINING TECHNICAL PROBLEMS

In the work to date, some technical problems have been identified whose solution would improve the performance of the technology. These technical problems include maldistribution of current among a number of bags operating off of one high voltage power supply, and the need for engineering an improved system for high voltage distribution from the power supply to the bags.

At the SWPS pilot unit, in which the resistivity of the fly ash was moderately high, it was noted that one or two of the bags would draw current while, at the same time, the remaining bags would not. In subsequent filtration cycles, the same bags might or might not have current. Back corona would bags, using the shake-deflate mechanism, the relative pressure drop would be low. With time, the pressure drop for the AESFF bags would continue rising until it would start to approach the pressure drop of the non-AESFF control bags. Low pressure drop for the AESFF bags could again be restored by vigorous cleaning. For the spreader-stoker pilot unit, which operated with low resistivity fly ash, the problems due to maldistribution of current and the resulting increase in pressure drop with time were noted.

Three general approaches to mitigating the current maldistribution and resulting pressure drop rise with time suggest themselves. The first is increasing the cleaning energy to the bags, which will keep them cleaner and consequently less susceptible to current maldistribution. The second approach is to condition the particle matter with an agent, such as SO_3, to lower the resistivity of the material. The third approach is to use a pulsed-energization powering arrangement to provide a more uniform current distribution [10]. Alternatively, the more economical intermittent energization type of power supply might also improve current distribution.

For a multiplicity of AESFF bags, operating off of one high voltage power supply, a simplified engineered approach for distribution of the electrical energization is desired. For both of the AESFF pilot units, individual electrical cables were brought to each of the bags. This approach was complex and led to electrical breakdowns that required a significant amount of maintenance.

CONCLUSION

A mathematical engineering design model has been developed that predicts the non-uniform particulate matter deposition and relative pressure drop, compared to conventional or non-electrified deposition for fabric filtration with integral particle charging and collection in a combined electric and flow field. The model prediction appears to be reasonably accurate on the basis of limited pilot plant data.

The mathematical engineering model which relates performance to electrical conditions, bag geometry, and particle size distribution is a design tool that can be useful in further development of this new technology.

The application of AESFF results in a decreased penetration of particulate matter through the filtration media as compared to conventional filtration. The decrease in penetration was achieved under electrical conditions in which the corona current was relatively low. With a higher and more optimum corona current, it is expected that the penetration of particulate matter through the filtration media will be decreased even further.

Remaining technical problems to be solved for the AESFF technology include providing a uniform distribution of current among multiple bags being operated by one high voltage power supply, and engineering a low cost and reliable means for energizing individual bags in a multiple bag installation.

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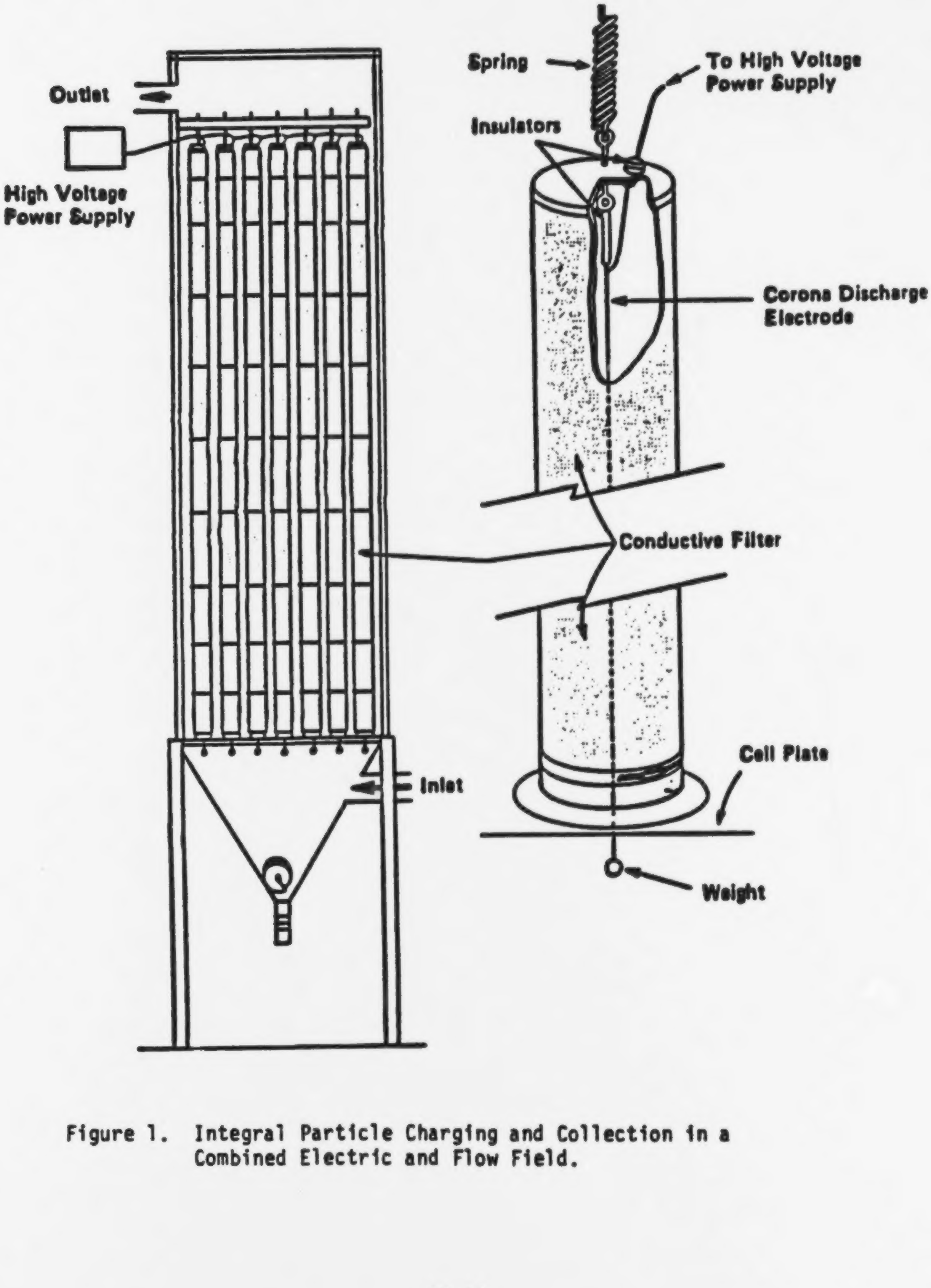
Outlet
High Voltage Power Supply
Inlet
Spring
To High Voltage Power Supply
Insulators
Corona Discharge Electrode
Conductive Filter
Cell Plate
Weight

Figure 1. Integral Particle Charging and Collection in a Combined Electric and Flow Field.

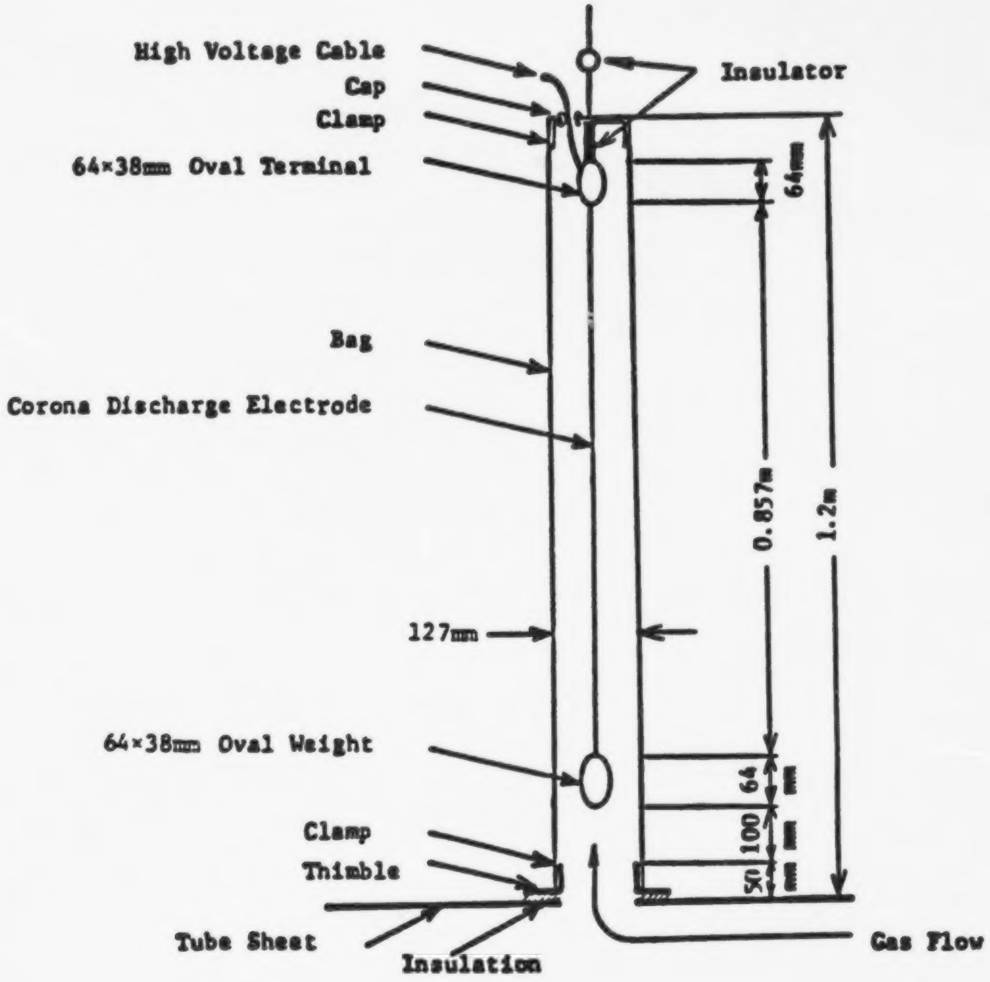

Figure 2. Single-Bag Experimental Baghouse Setup.

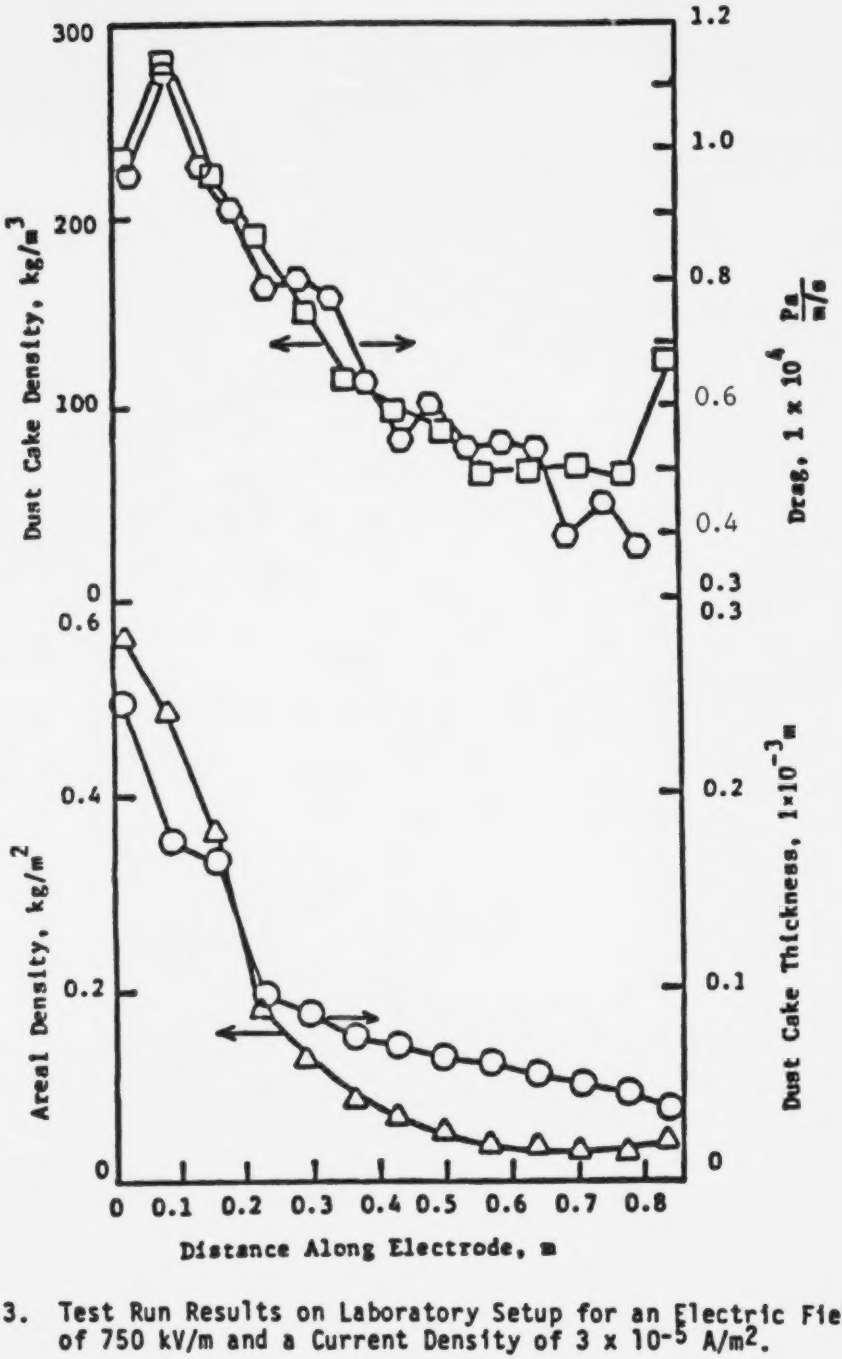

Figure 3. Test Run Results on Laboratory Setup for an Electric Field of 750 kV/m and a Current Density of 3 x 10^{-5} A/m^2.

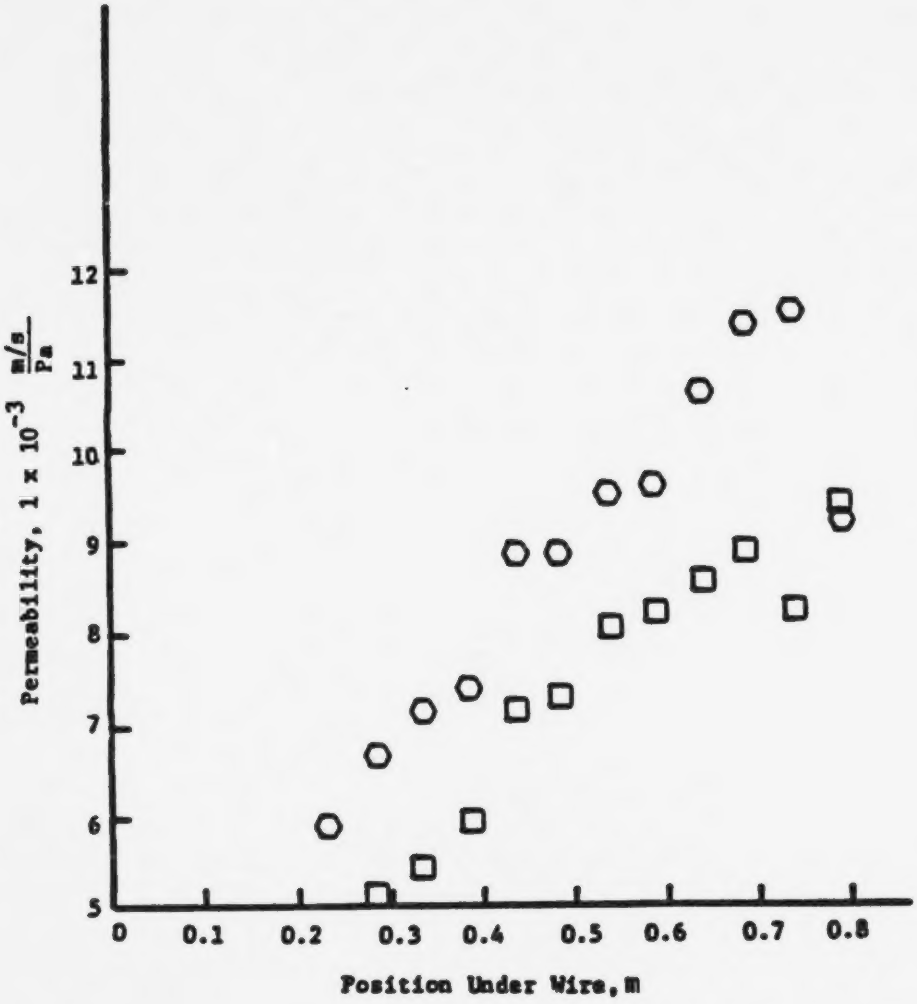

Figure 4. Examples of Linearity of Face Velocity at Unit Pressure vs. Position Under Corona Wire.

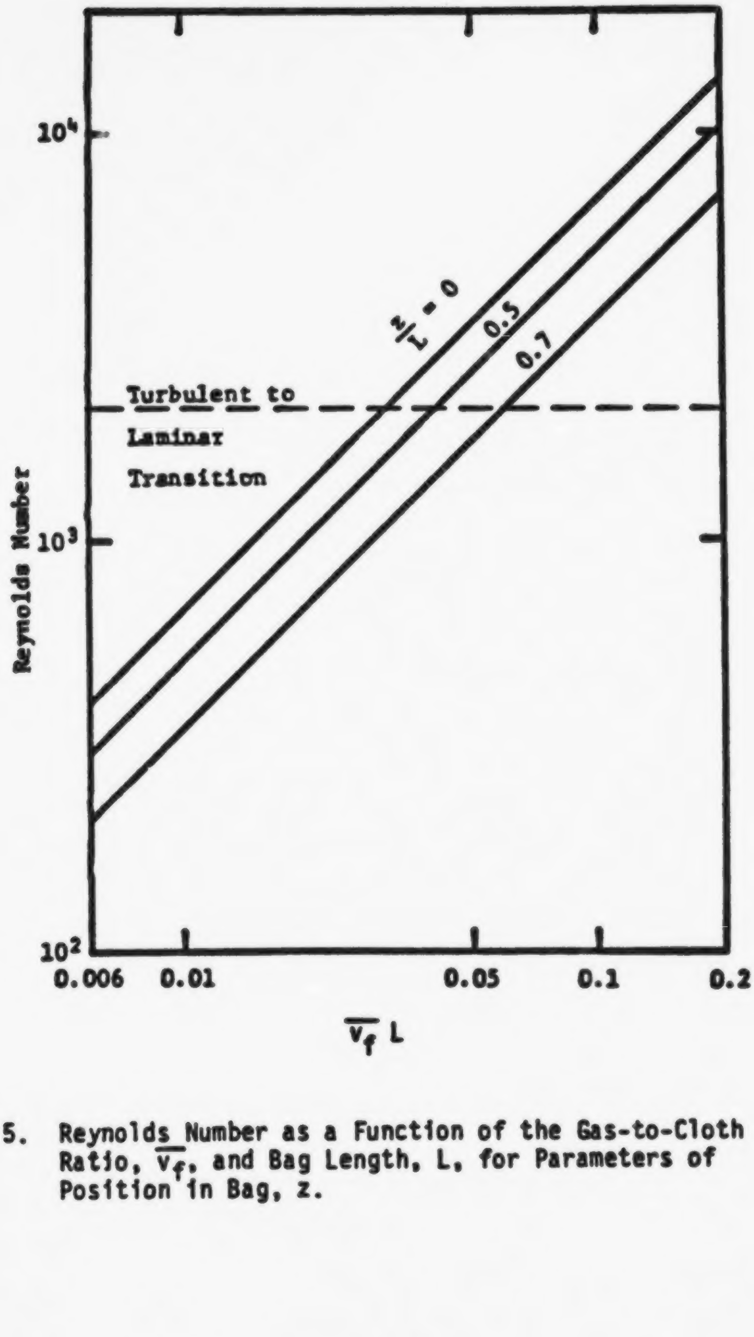

Figure 5. Reynolds Number as a Function of the Gas-to-Cloth Ratio, $\overline{v_f}$, and Bag Length, L, for Parameters of Position in Bag, z.

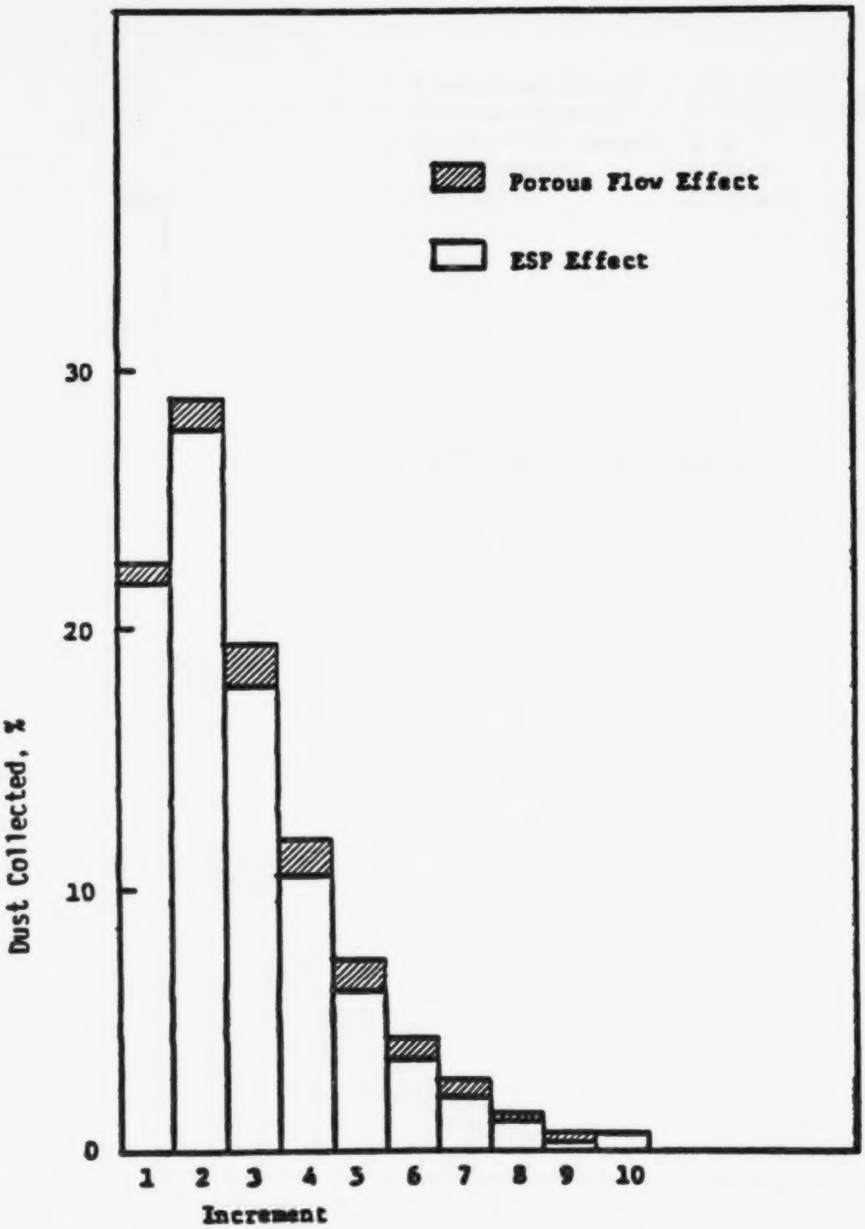

Figure 6. Mathematical Model Prediction of Non-uniform Distribution for an Electric Field of 750 kV/m and a Current Density of 3×10^{-5} A/m^2.

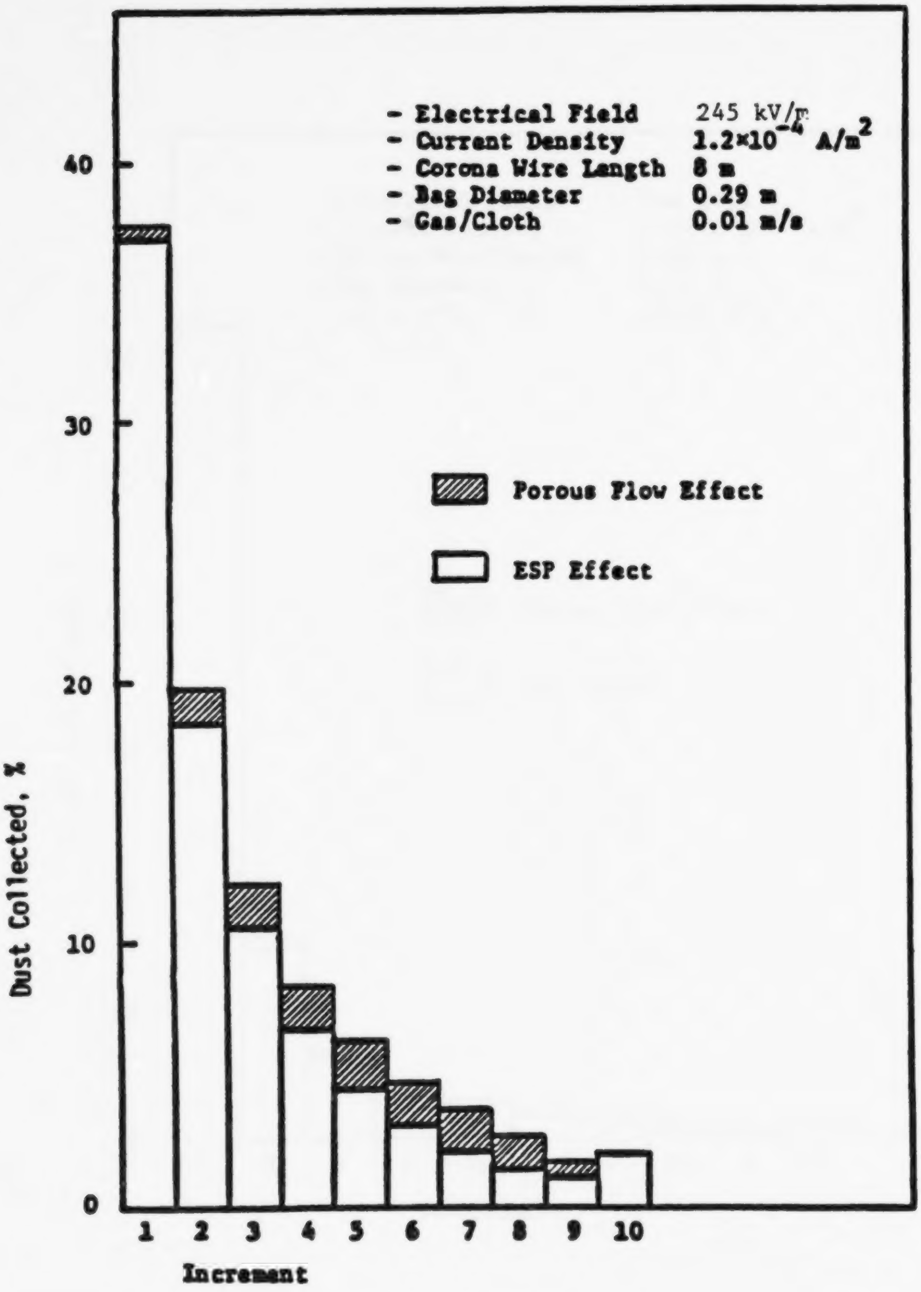

Figure 7. Mathematical Prediction of the Non-uniform Distribution Using the SWPS Pilot Unit Electrical Conditions.

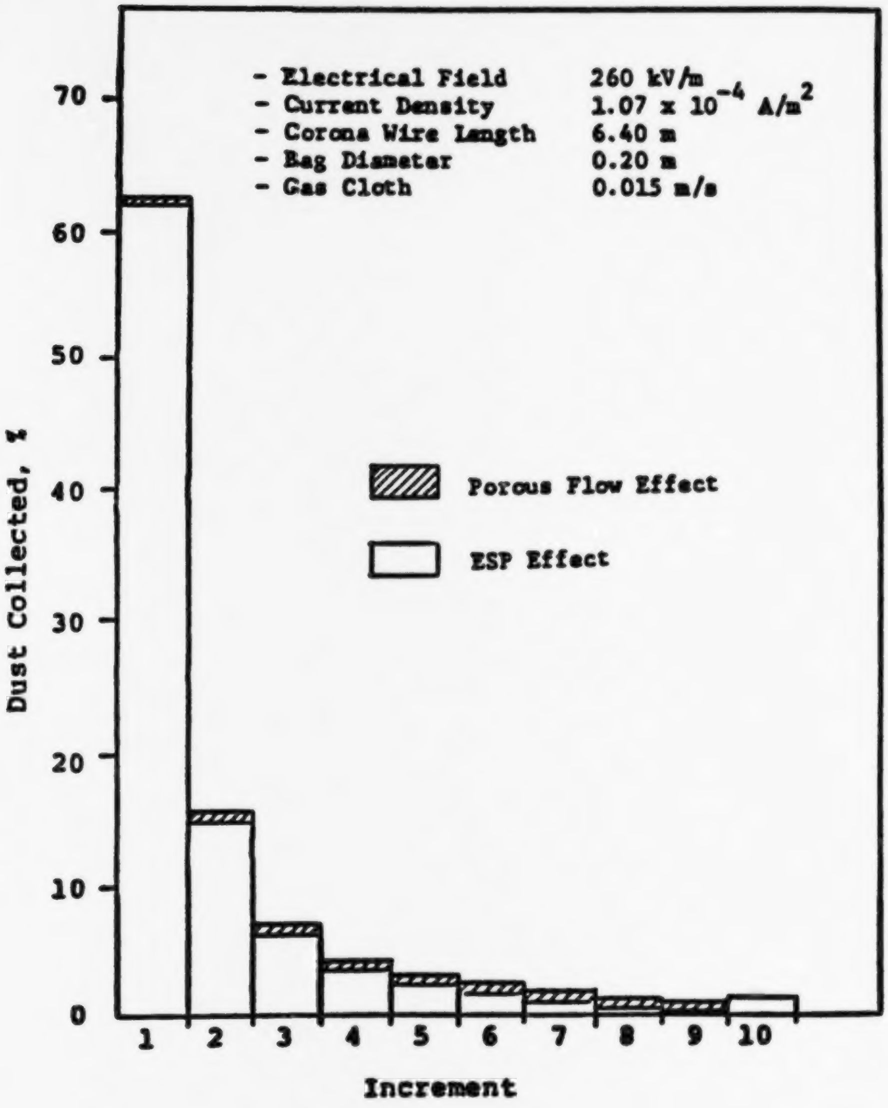

Figure 8. Mathematical Model Prediction of the Non-uniform Deposition for the Spreader-Stoker Pilot Unit Electrical Conditions.

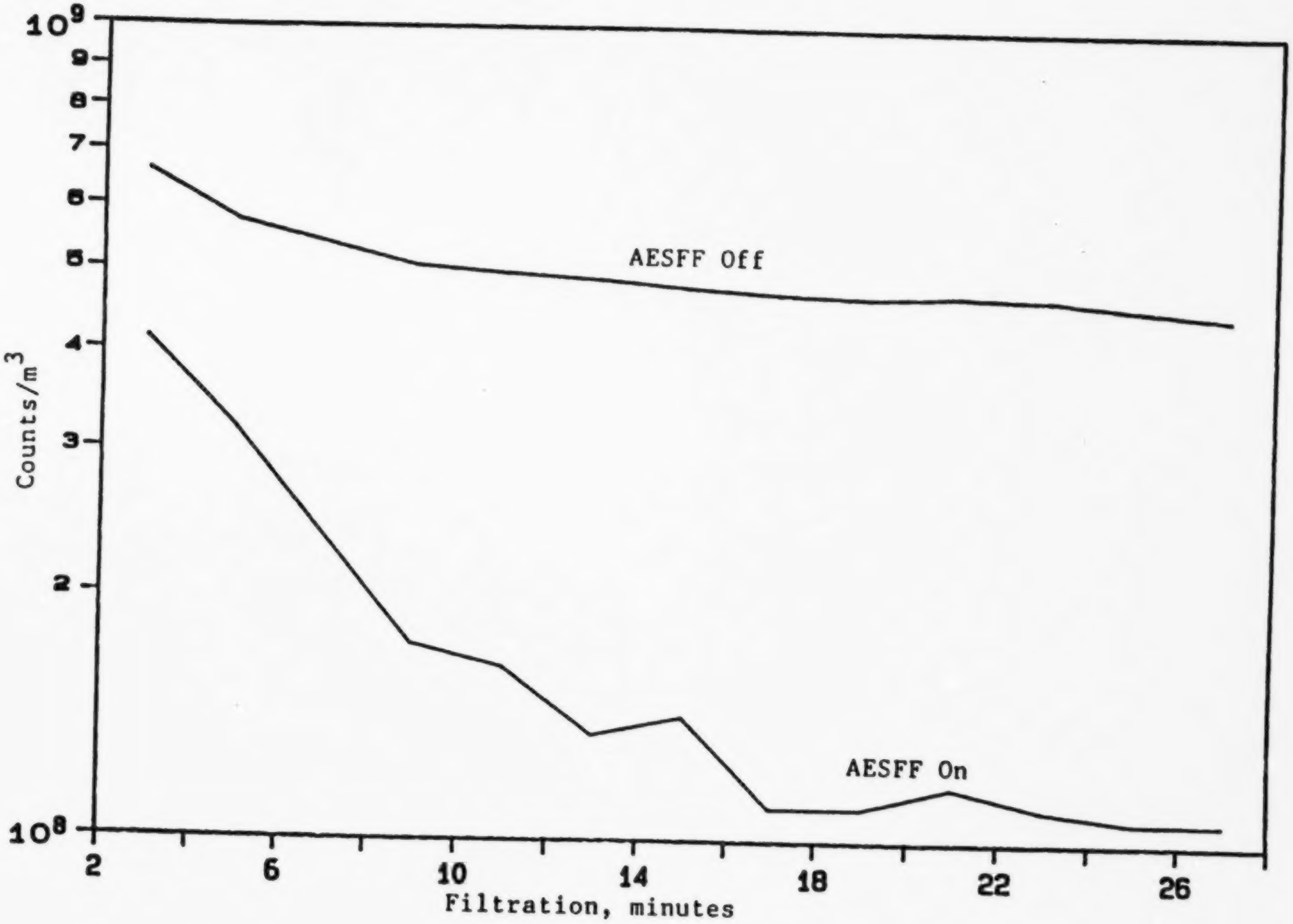

Figure 9. Particle count in exiting gas stream with and without AESFF--particle size range, 0.3-1.0 μm .

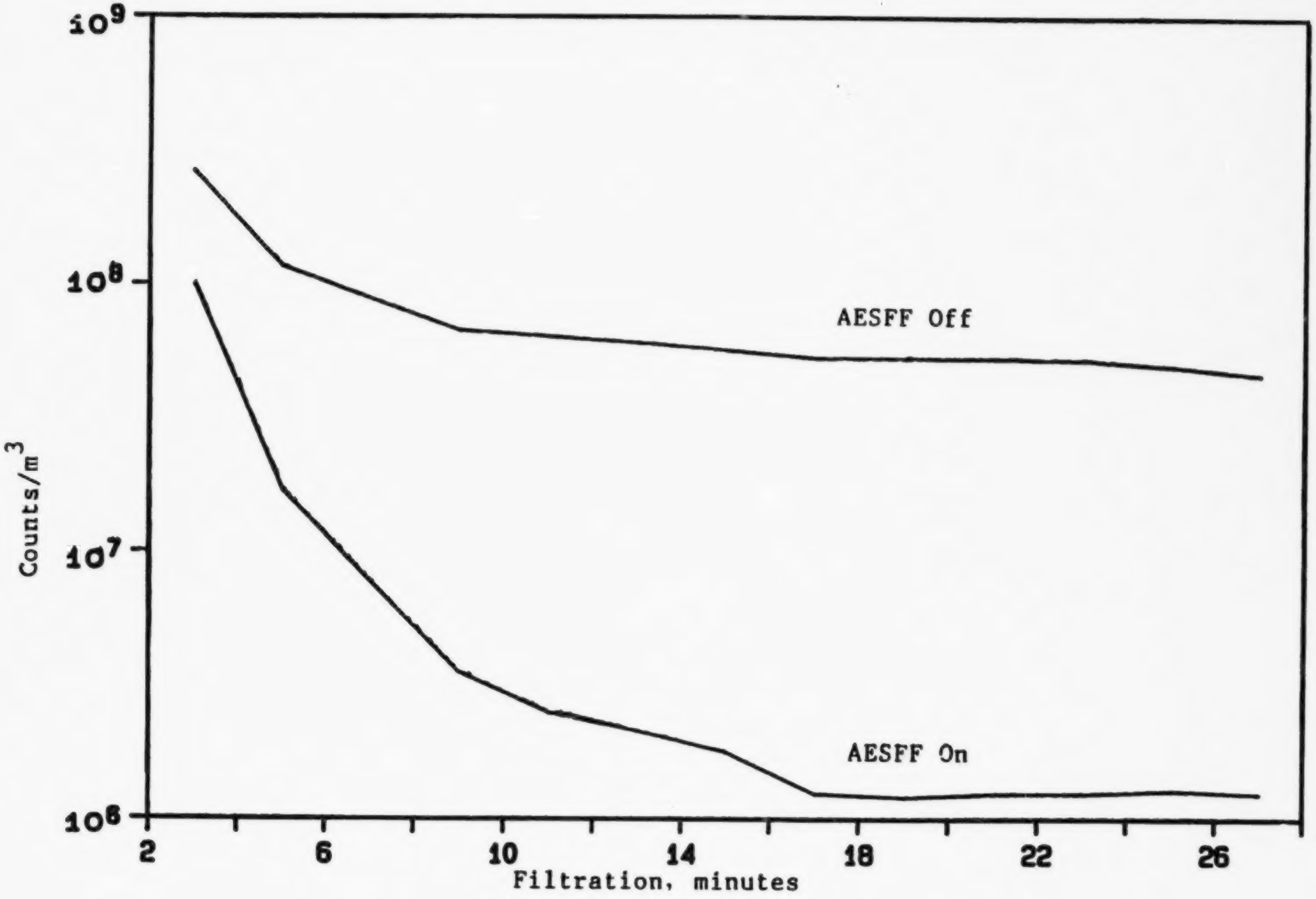

Figure 10. Particle count in exiting gas stream with and without AESFF--particle size range, 1.0-3.0 μm

USING THE FABRIC FILTER GAS FLOW MODEL TO SIMULATE THE OPERATION OF A PILOT-SCALE BAGHOUSE

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ABSTRACT

A new microcomputer-based model for simulating gas flow in a baghouse has been developed. It is designed to calculate the various gas flow rates through the manifold paths of a multichambered fabric filter. It accounts for the time-varying drag of the individual compartments as well as for the aerodynamic drag of the ducting and the input and output losses for each compartment. This paper describes the use of this model to investigate the effects of varying operating conditions or parameters on the pressure drop across the system. The model was used to simulate the behavior of a pilot-scale baghouse for the examples treated in this paper. The pilot plant is useful for this demonstration because its configuration is relatively simple and its behavior has been thoroughly characterized under a range of operating conditions.

INTRODUCTION

In a conventional utility baghouse, the bags are cleaned by briefly reversing the flow of gas through them or by gently shaking them by their suspensions. Both methods require temporarily removing bags from service for cleaning. The usual way to accommodate off-line cleaning is to divide the baghouse into a number of compartments and use a rotating schedule for cleaning the bags. This mode of operation produces a condition in which no two compartments are in the same phase of their filtration cycle at any given time. Consequently, the thickness of the dust layer on the fabric and the resultant flow resistance differs from one compartment to another. In general, the compartment most recently cleaned has the lowest flow resistance and therefore accepts more gas flow than does any other compartment (1).

Models of baghouse operation begin with the relationship between the pressure drop across a unit area of filter material and the rate of gas flow through it. If the flow rate is held constant during filtration, the pressure drop increases because of the additional dust deposited on the dustcake. The increase is not typically linear, because the dustcake is irregular, cracked, and broken after undergoing the cleaning phase (2,3). The drag (ratio of pressure drop to flow rate per unit area) normally rises sharply at first, as shown in figure 1. This rapid change occurs as the cracks and crevices in the dustcake fill with fresh dust (4).

As the surface of the dustcake evens out, the rate of increase slows down and becomes more linear.

In a single compartment of a baghouse, it is normally assumed that the distributed effects of irregularities in the dustcake can be represented by a typical unit of area, since all of the bags in a given compartment are treated to a common schedule of filtration and cleaning. The problem of describing the distribution of gas flow through a compartmented baghouse is complex, however, because the flow resistance of any one compartment is affected by the flow resistance of all of the others. The complexity of the problem is compounded by the pressure losses in the interconnective ducting.

A number of useful baghouse models have been developed, but in most of these models the pressure drop in the connecting ducts has been regarded as negligible (5,6,7). In a recent paper, Muir and Ahmadzadeh (8) described a model that includes inlet and outlet pressure losses for individual compartments, but those of the ducts that tie the system together are not modeled.

The present model is based on an iterative sequence of quasistatic calculations of flow rate and tubesheet pressure drop for all compartments, and it includes not only the aerodynamic drag of the fabric and dustcake, but also that of the connective ducting, dampers, and valves. The total gas flow through the baghouse is held constant throughout the simulation. Results of some calculations are also shown to indicate the kinds of applications that can be treated with this model.

GENERAL CONFIGURATION OF THE MODEL

The baghouse model can simulate either of the two general arrangements of compartments shown in figure 2, as well as those that can be derived from these two configurations by simple extensions and deletions. Since each segment of duct has physical parameters and a flow rate that may differ from those of the other segments, the pattern of flow resistances that must be set up for the calculations is complex. Figure 3 shows the generalized flow pattern and the relationship of resistances, each of which is represented by a rectangle. $R_B(i)$ represents the resistance of the dustcake and fabric. It is the only resistance that is not a quadratic function of the gas velocity, and the only one that varies with time. In the present version of the model, $R_B(i)$ is computed from the effective residual drag S_E and the normalized drag K_2 as expressed in the equation

$$\Delta P = S_E V + K_2 V W \tag{1}$$

in which V is the average value of the component of the velocity normal to the surface of the fabric, and W is the areal density of the dustcake minus the residual part (9).

The resistances $R_A(i)$ represent the effects of the inlet opening of a compartment, the hopper, the tubesheet and any other components or structures between the inlet and the surface of the dustcake. $R_C(i)$ accounts for the outlet resistances (e.g. outlet poppet valves) from the surface on the clean side of the fabric to the outlet duct of the compartment. $R_D(i)$ and $R_E(i)$ are the resistances of the connecting ducts.

Other than the flow resistances described in the last paragraph, the total flow rate Q is the only parameter that is maintained at a specified constant value for this model. The values of gas flow rate and pressure drop for the various compartments are all interactive. Any change that modifies the gas flow through one compartment will affect the flow in all compartments. For example, if a compartment is taken off line for cleaning, the gas flow rate in all other compartments must increase to maintain Q at a constant value.

The model will accommodate up to 50 compartments, any number of which may be designated as off line for a particular simulation. The sequence for compartment cleaning can be set up in any of four different ways. Two modes are based on batch cleaning: one is initiated by an input value of pressure drop and the other is set to begin at a given time after the previous cleaning. The third mode is to clean the compartments in a sequence of uniform time intervals over the duration of the complete operating cycle of the baghouse. Finally, the fourth option allows the operator to assign a specific time in the cycle for each compartment to be cleaned.

COMPUTATIONAL METHOD

Since the quantities involved in the model include linear, quadratic, and time-dependent terms, an approach based on a closed solution of a general set of equations does not appear to be feasible. An applicable method commonly used for solving fluid network problems is based on a series of successive approximations (10). First, a flow rate is assumed for every branch of the system so that continuity of mass flow is conserved at every junction. A correction for each flow rate is then calculated, applied, and recalculated until the desired degree of accuracy is achieved. This method was tried during the development of the model, but the convergence to an acceptable degree of accuracy was too slow to be practical. Such a general approach can be improved upon, however, for the specific types of flow networks that apply to baghouses. The method used in this model requires assuming a value for only one branch of the network and finding corrections to that quantity leading to a self-consistent solution.

The solution is based on a fixed value of the total gas flow rate through the baghouse. Computation begins with the assumption of a flow rate for only one compartment, namely, the one farthest from the inlet. The values of the flow rates and the corresponding pressure drops in all other branches are then computed one at a time, working from the specified outermost branch back toward the inlet. If the calculated value of the total gas flow does not agree with the given value, then the assumed flow rate through the outermost compartment is adjusted proportionally to the error, and the calculation is repeated. Convergence to an accuracy of about 0.1% typically occurs within no more than five to 20 iterations, depending upon the number of compartments there are in the simulated baghouse.

What has been described so far is a method for finding static solutions for pressure drop and flow rate. A sequence of such states simulates the dynamic behavior of the baghouse. During each increment of time dust is added to the fabric in each compartment, increasing the dustcake resistance. The amount of dust added in each compartment is the product of the mass loading and the gas flow rate for that compartment.

Before a simulation can be run, a variety of input data must be provided. Provisional values for these data are kept in a separate file, and an auxiliary program is available for resetting the values for a specific application. Thus, it is not necessary to reset all input data at the start of each simulation. Information required to set the configuration and general operation includes

- number of compartments
- total gas flow rate
- inlet mass loading
- bag diameter
- bag length
- duration of cycle
- reverse-gas A/C

These items are self-explanatory. A second category of parameter is one common to all compartments, including:

- effective residual drag (S_E)
- normalized drag (K_2)
- number of bags per compartment

The quantities S_E and K_2 may be taken from empirical studies on pilot-scale work. When more accurate methods are developed for predicting S_E and K_2 from coal or ash chemistry, auxiliary programs or modifications to the flow model can be provided.

The flow model also allows for setting separate values of S_E, K_2 and number of bags for individual compartments. It is also possible to designate any number of compartments off line throughout the simulation.

In addition to the above data, the compartment cleaning schedule is set up before starting the simulation. Cleaning can be by means of a timed mode or a batch mode. If the latter is selected, it may be triggered by flange-to-flange pressure drop or by elapsed time.

Finally, the drag coefficients for each compartment and its ducting must be specified by reference to the flow resistance components illustrated in figure 3, except for the $R_B(i)$, which have already been given in terms of S_E and K_2. In well-defined, elementary structures, such as straight runs of ducting, the drag coefficients may be calculated. For more complex configurations, the coefficients can be estimated from similar existing systems, or they can be derived from scale model tests.

The following quantities are computed for each increment of time:

- tube sheet pressure drop for each compartment
- flange-to-flange pressure drop
- gas flow rate for each compartment
- inlet and outlet pressure drop for each compartment
- pressure drop across each identifiable segment of duct

All of these quantities are accessible for output; however, the model in its present configuration provides for two output modes. One is a graphic display of the flange-to-flange pressure drop and the tubesheet pressure drop of a specified compartment plotted as functions of the elapsed time. The other mode is a table of the same data for each time increment plus the gas flow rate through the given compartment.

COMPUTED RESULTS

The baghouse model was used to simulate the operation of EPRI's four-compartment pilot-scale baghouse at the Arapahoe Steam Plant in Denver, Colorado. This baghouse is an ideal subject for modelling because it has been characterized under a relatively broad range of operating conditions. The matrix of tests included in this study included a range of air-to-cloth ratio (A/C) between 1.0 and 3.5 ft^2/acfm, operating cycles of 1, 2, and 3 hours, and two levels of boiler load. Reverse-gas cleaning is simulated.

Figure 4 is a typical plot derived from the model. It shows the computed values of the flange-to-flange pressure drop and the tubesheet pressure drop for compartment 1. The length of a complete operating cycle is two hours, and the compartments are cleaned according to a timed schedule of equal intervals between the cleaning of successive compartments. At the beginning of the simulation, the baghouse is brought on line with all compartments cleaned down to the residual dustcake, according to the most common practice. When any compartment is taken off line for cleaning, there is a sharp increase in the flange-to-flange pressure drop and in the tubesheet pressure drop for all other compartments. When the cleaned compartment is brought back on line it accepts a greater fraction of the gas flow than it did before cleaning, so the tubesheet pressure drop for all of the other compartments drops below its previous value. In figure 4, the difference between the magnitude of the flange-to-flange pressure drop and the tubesheet pressure drop for compartment one arises from the simulated pressure losses in the ducting.

The solid lines in figures 5 and 6 were generated by the computer simulation for the range of conditions indicated. The data points shown are the results of measurements made at the pilot plant. As a result of the tests, values for S_E and K_2 were determined. Each of these quantities was found to be well represented by a linear function of the air-to-cloth ratio, or face velocity, V. For low load (4.0×10^5 lb/h) the best fit is

- $S_E = 1.01 + 0.122\ V$
- $K_2 = 10.0 + 1.97\ V$

and for high load (8.0×10^5 lb/h),

- $S_E = 1.06 + 1.22\ V$
- $K_2 = 10.63 + 3.24\ V$

The curves representing average pressure drop as a function of air-to-cloth ratio are quadratic in form because of the combined effects of the gas velocity and the rate of particle deposition, each of which operates linearly on the pressure drop. Thus, if the amount of dustcake on a filter remains constant, then the pressure drop across it is directly proportional to the face velocity. In practice, however, increasing the face velocity produces an increase in the rate of growth in the dustcake density.

CONCLUSIONS

A new microcomputer-based baghouse model has been developed. It includes provisions for calculating the pressure drop in any segment of ducting, as well as in the individual compartments. The model is designed to be operated from built-in sequences of directions by an operator with a minimum of instruction.

The basic calculations and structure of model have been verified by comparison of model results with data taken at the fabric filter pilot plant. The unique aspect of the model is its treatment of the gas flow through the connective ducts, but because of a lack of data, the accuracy of that feature requires further verification.

The purpose of the model is to assist in the optimization of baghouse design and operation. By varying any of the input parameters, such as the number and design of compartments, and observing the effect on the average or peak value of pressure drop, one can determine which design is more cost effective. Similarly, the most effective cleaning strategy can be developed by observing the trends as the parameters are adjusted.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

ACKNOWLEDGMENTS

The work leading to the developments reported in this paper was supported by the Electric Power Research Institute, Palo Alto, California, within the scope of project number RP 1129-8, under the management and direction of Mr. R. C. Carr and Mr. W. V. Piulle. The authors also acknowledge the assistance of Dr. J. R. McDonald in the preliminary development of an approach to the model.

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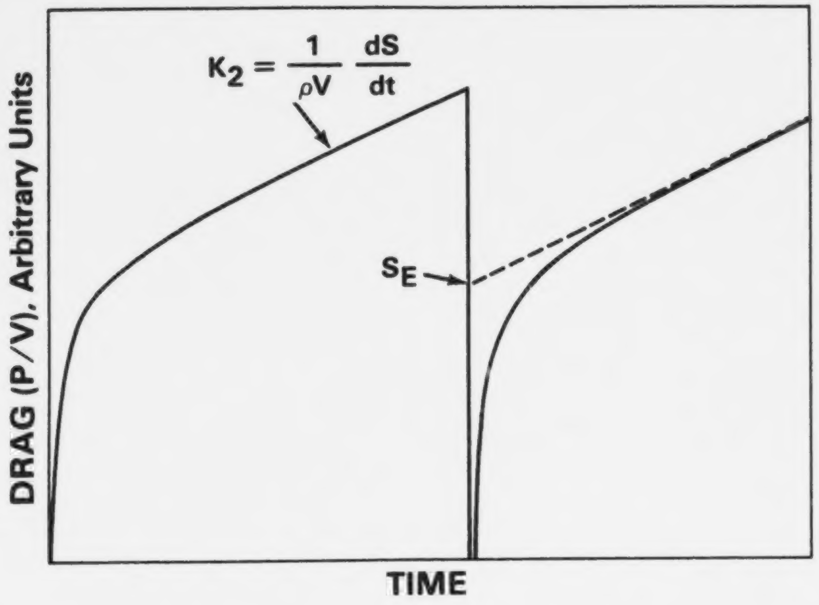

Figure 1. Cyclic filtration parameters.

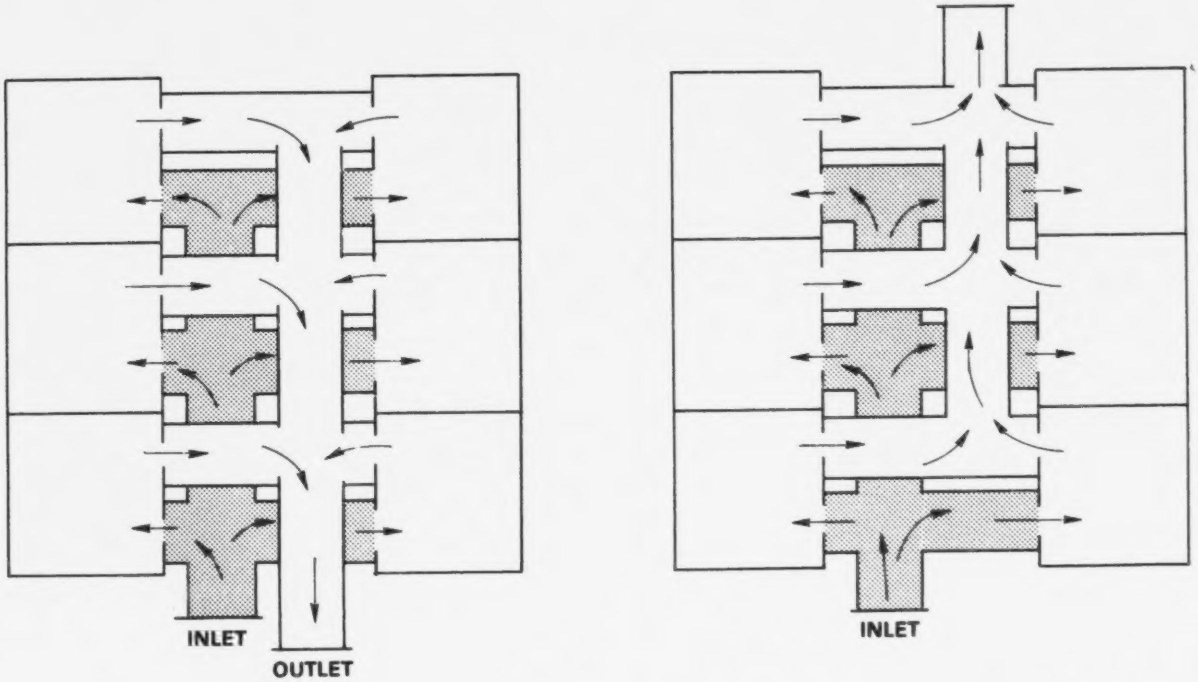

Figure 2. Baghouse configurations.

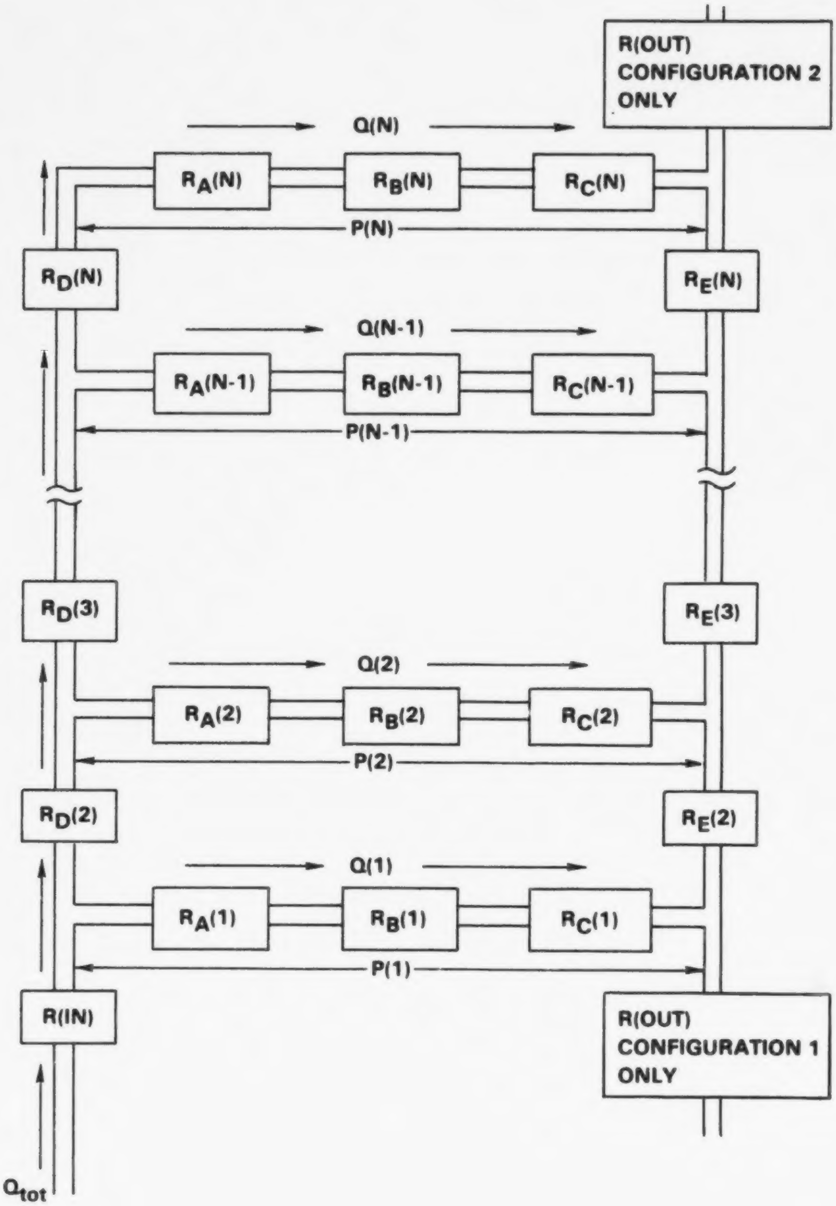

Figure 3. Flow resistance parameters.

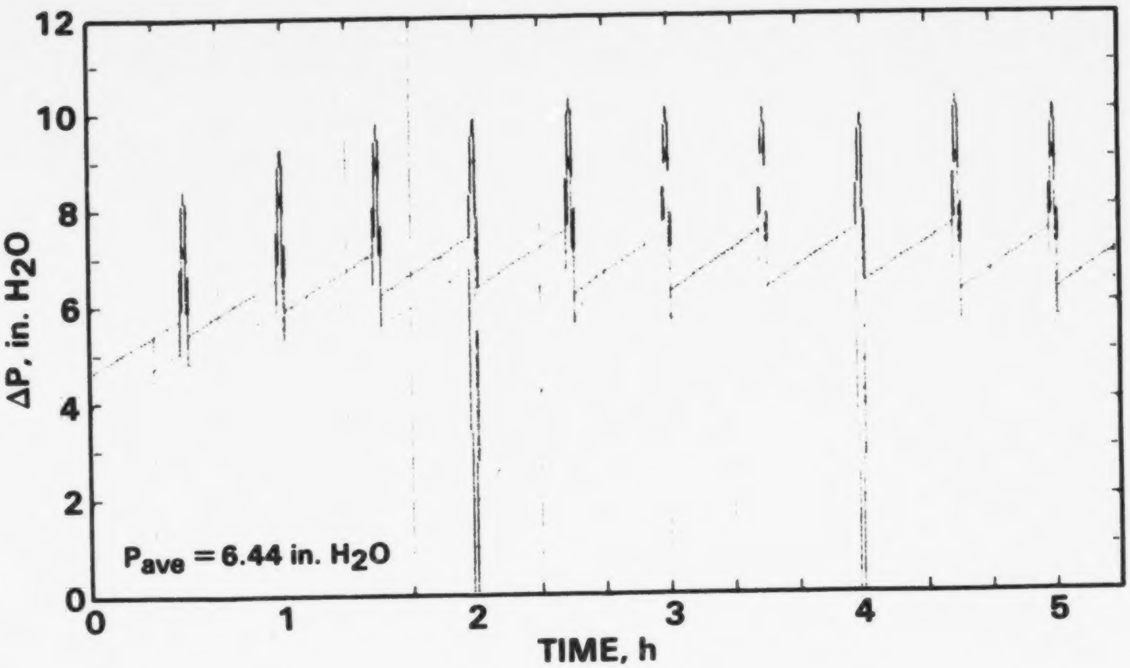

Figure 4. Typical graphic output.

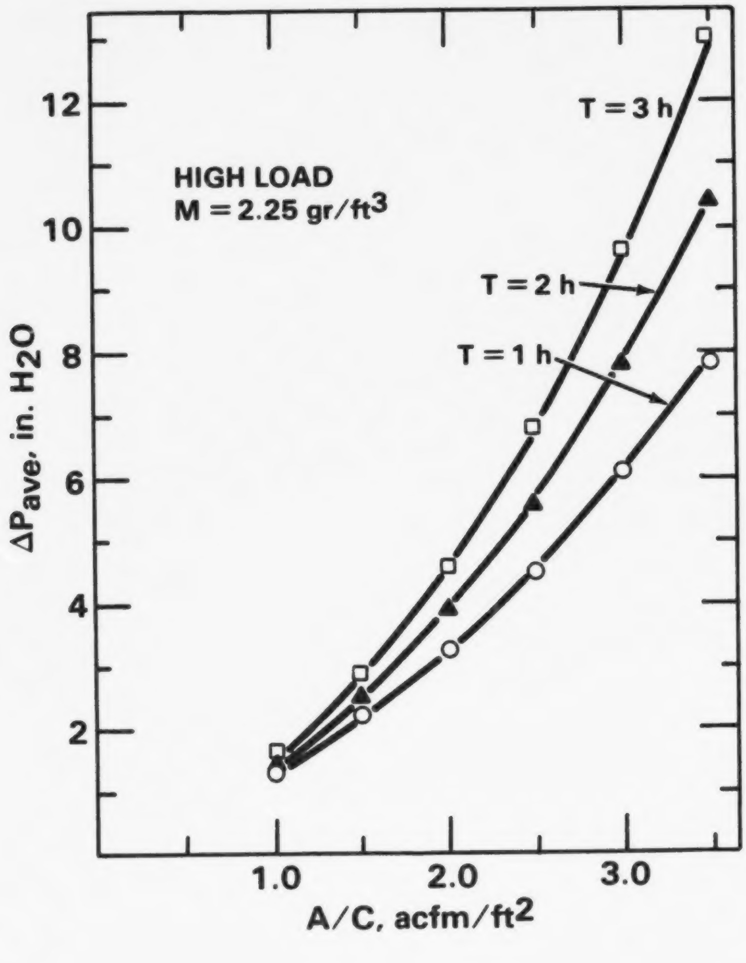

Figure 5. Comparison of model with pilot-scale results; high load.

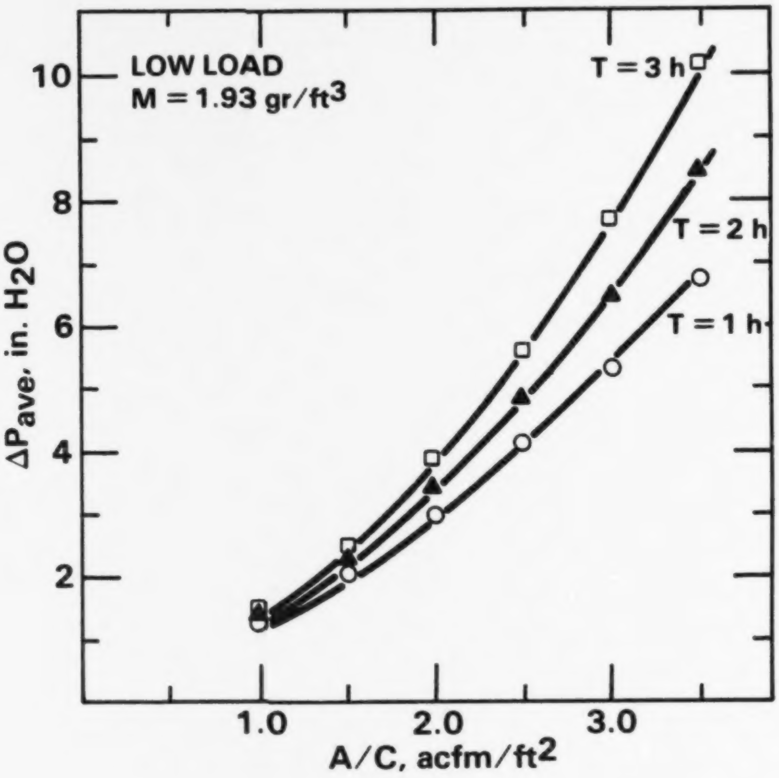

Figure 6. Comparison of model with pilot-scale results; low load.

FABRIC FILTER OPTIONS FOR LOW-SULFUR COAL

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ABSTRACT

Parametric studies of fabric filter performance collecting fly ash from low-sulfur coal boilers have been in progress since 1980 at EPRI's Fabric Filter Pilot Plant (FFPP) located at the Arapahoe Test Facility in Denver, Colorado. These tests have been conducted to explore and evaluate the options available to the utility industry in implementing fabric filtration for particulate control. Parameters studied have included fabric construction, filtering air-to-cloth ratio, filtering time, and details of several bag cleaning methods. Cleaning methods studied include reverse gas, reverse gas with sonic assistance, and shake/deflate. As a result of this research, families of curves were developed showing the optimum range of operation of baghouses with each bag cleaning method and for several fabric constructions. This work was also coordinated with concurrent supporting research reported in companion papers. The results of the parametric tests provided input for recent operation of the FFPP in which a demonstration of improved or "state of the art" baghouse operation was conducted. Each of the four FFPP compartments was set up for a different cleaning method. These were reverse gas, reverse gas with intermittent sonic assistance, reverse gas with sonic assistance, and shake/deflate cleaning. The operating parameters for each compartment (filtering air-to-cloth ratio, filtering time, and cleaning parameters) were selected to demonstrate advanced baghouse operation and typically exceeded current utility practice. During the 5000 hours of operation, drag, residual dustcake weight, and emissions were measured for each compartment. Data developed during this test confirmed that drag associated with shake/deflate cleaning and reverse-gas cleaning with sonic assistance was about 30% lower than for reverse-gas cleaning alone and that residual dustcake weights were about 50% lower for these more energetic cleaning methods. Emissions were significantly higher for shake/deflate cleaning and reverse-gas cleaning with sonic assistance, ranging from 1.5×10^{-4} to 1.5×10^{-3} lb/10^6 Btu, but were still an order of magnitude less than NSPS regulations for new stationary sources (0.03 lb/10^6 Btu).

INTRODUCTION

The use of fabric filter baghouses for particulate emission control from coal-fired power plants has rapidly gained acceptance within the utility industry. This is demonstrated by the extensive growth in utility commitments to baghouses, increasing from 0 to 20,000 MW(e) capacity within the past ten years.

The utility experience with baghouses has been very good overall, especially on western subbituminous coals. The outstanding particle collection efficiency and "clear stacks" corresponding to invisible emissions experienced by almost all well-maintained units are primarily responsible for their popularity, although economic studies indicate that there are frequently cost advantages to be realized over alternative control systems as well. Also, research indicates that future technical developments will reduce capital and operating costs even further.

To support the utilities in their application of fabric filtration for air pollution control, the Electric Power Research Institute in 1980 began a research program to evaluate the options available with current baghouse technology. The program has included pilot-scale, full-scale, and laboratory studies. The early work dealt with baghouses cleaned by reverse gas because 90% of full-scale utility baghouses at the time were cleaned by that method. To characterize other fabric filter options available to the utility industry, the program, more recently, has concentrated on the evaluation of reverse-gas cleaning with sonic assistance, shake/deflate cleaning, dustcake properties, and the performance of different fabrics. The results of these studies have recently been applied to a demonstration of improved baghouse operation in which the most favorable operating parameters have been evaluated during a year-long test. This paper presents an overview and summarizes key results from 6 years of parametric evaluations of fabric filter operation and the results of a test to demonstrate low-sulfur coal baghouse operation with optimized operating parameters for four bag cleaning methods. The testing took place at a 10 MW(e) Fabric Filter Pilot Plant (FFPP) located at EPRI's Arapahoe Test Facility in Denver, Colorado. This pilot-plant baghouse takes flue gas from a slip-stream from a 110-MW coal-fired boiler burning western, subbituminous, low-sulfur coal (0.7%). Also, at this facility is a 2.5 MW(e) pilot-scale baghouse, where additional testing on different fabrics was conducted. Both of these pilot units are instrumented for continuous analysis. Gas flow to the pilot plants is provided by induced draft fans on each system. This allows control of the filtering air-to-cloth ratio independently of boiler load. Full-sized bags and hardware are used to simulate performance of commercial systems. In order to extend the database and verify the pilot-scale results, data were collected from 12 full-scale utility baghouses. Laboratory studies of reverse-gas cleaning with sonic assistance provided additional supporting data.

During the first six years of operation of the Fabric Filter Pilot Plant parametric evaluations of reverse-gas cleaning, reverse-gas cleaning with sonic assistance, and shake/deflate cleaning took place. Operating parameters evaluated included reverse-gas air-to-cloth ratio, sound pressure and power-weighted mean frequency of sonic horns, duration of horn sounding and frequency of horn use, shake amplitude and shake frequency, shake duration, and deflation air-to-cloth ratio and duration. In addition, pressure drop was evaluated for various filtering air-to-cloth ratios and filtering times. Performance parameters included drag, emissions, and residual dustcake areal densities. In a separate test, fabrics with different levels of exposed surface texturization were evaluated. As a result of these parametric tests it was possible to formulate a set of operating parameters to allow a demonstration during the past year of improved or "state of the art" baghouse operation.

TECHNICAL DISCUSSION OF THE PARAMETRIC TESTS

REVERSE-GAS CLEANING

Reverse-gas cleaning was the first cleaning method evaluated at the FFPP. This cleaning method was determined to be inadequate in controlling pressure drop and residual dustcake weight. Originally, the reverse-gas air-to-cloth ratio had been set within the range typical of full-scale utility baghouses, 1.5 to 2.0 acfm/ft^2. With pressure drop initiated cleaning, tubesheet pressure drop increased until a mode of continuous cleaning was occurring. This led to an evaluation of time-initiated cleaning with time intervals between compartment cleanings set at 60, 90, 120 or 360 minutes. Although baghouse operation could be better controlled with this cleaning initiation scheme, drag and bag weights were still high and long-term increases in these parameters could not be controlled. This led to a parametric study of the reverse-gas air-to-cloth ratio. Reverse-gas air-to-cloth ratios from 1 to 4.5 acfm/ft^2 were tested. Average tubesheet pressure drop was measured at each reverse-gas setting. The test results are presented graphically in figure 1. Also shown in this figure is the range of reverse-gas values usually found at baghouses used in the utility industry. As can be seen, the average pressure drop was not a strong function of the reverse-gas air-to-cloth ratio. The experience with reverse-gas cleaning accelerated the implementation of test programs to evaluate more energetic baghouse cleaning techniques.

REVERSE-GAS CLEANING WITH SONIC ASSISTANCE

In tests conducted at the Fabric Filter Pilot Plant, four parameters specific to the application of sonic assistance to reverse-gas cleaning were evaluated. These included the average sound pressure created in the compartment by the horn, the frequency of the horn, the duration of horn sounding, and the frequency of horn use. Supporting measurements were conducted in the laboratory to further investigate dustcake removal from fabric samples as a function of power-weighted mean frequency, sound pressure, and duration of horn sounding.

Sound Pressure

Measurements in the laboratory and at full-scale baghouses were conducted to determine the dependence of residual dustcake removal on localized sound pressure. In the laboratory, fabric/dustcake samples from the FFPP were subjected to various sound pressure values at three different frequencies. The percentage of original dustcake removed after a 30-s exposure was measured. The data are presented in figure 2. Sound pressures of at least 80 Pa are necessary to dislodge a significant fraction of the dustcake for frequencies of 100 or 200 Hz. Similar results were obtained in full-scale baghouse tests where horns with a frequency of 211 Hz were in use. These data are shown in figure 3. Again, significant dustcake removal requires local sound pressure of 80 Pa or greater. For comparison data from baghouses filtering fly ash from utility boilers burning high-sulfur eastern coal are shown. Significantly higher sound pressures are required for adequate dustcake removal.

Horn Frequency

Experiments conducted at the FFPP and in the laboratory have demonstrated the impact of horn frequency on pressure drop reduction. In this paper the power-weighted mean frequency is used to describe horn frequency. It is the logarithmic average of the frequency spectrum, weighted by the power at each frequency. In the laboratory, fabric/dustcake samples removed from the FFPP were subjected to 30-s exposures of sound pressure at four different frequencies and three sound pressures. The percentage of dustcake dislodged during the 30-s exposure was measured. The data are presented in figure 4. Significant dustcake removal occurs with frequencies below about 250 Hz. These data also confirm the data reported above concerning the necessity of high sound pressures to insure adequate dustcake removal. The effect of frequency on tubesheet pressure drop during reverse-gas cleaning with sonic assistance is shown in figure 5. These data were developed at the FFPP. Again, frequencies below about 250 Hz are best for reducing pressure drop and dustcake.

Duration of Horn Sounding

Measurements were conducted in the laboratory to determine the optimum duration for horn sounding. Fabric/dustcake samples from the FFPP were subjected to five different sound pressures for cumulative periods of 10, 30, 60, and 90 s. The frequency was 200 Hz. The data are shown in figure 6. Dustcake removal is essentially complete after 30 seconds of horn actuation. This time period is typical of the length of reverse gas at most full-scale baghouses and indicates that the simultaneous sounding of the horns during reverse-gas cleaning is practical. Other experiments at the FFPP demonstrated that horn sounding during filtering or in the absence of reverse gas is ineffectual in dustcake removal.

Frequency of Horn Use

A pilot-plant test program was conducted to determine the optimum frequency of use for sonic horns. Horn actuation was varied from once every third day up to once during every cleaning sequence (eight times per day). The lowest operating pressure drop was achieved when the horns were used every time the bags were cleaned, although significant benefits were observed when the horns were used as infrequently as once a day. These data are shown in figure 7. Since there is no indication that bag deterioration occurs because of horn use and since the power consumption by the horns is very low, it would obviously be most beneficial to use horns during each cleaning sequence.

SHAKE/DEFLATE CLEANING

During the evaluation of shake/deflate cleaning several parameters were investigated. These included shake amplitude, shake frequency, duration of shaking, deflation air-to-cloth ratio, and duration of deflation.

Shake Amplitude and Frequency (Bag Cap Acceleration)

With shake/deflate cleaning, the energy imparted to the bag to dislodge the dustcake can be characterized by the amplitude of shaking and the frequency of shaking. At the FFPP, parametric tests were performed to study the optimum shake amplitude and shake frequency consistent with the lowest energy input to the bags, the lowest drag and most effective dustcake removal. The energy can be characterized by the acceleration imparted to the bag cap, and it is represented by the following equation

$$a = 4(\pi)^2 f^2 Ak,$$

where

a = acceleration imparted to the bag, g,
f = frequency of shaking, Hz,
A = amplitude of shaking, inches,

k = empirical constant, 0.75 for our work, and
1g = 386 in./s^2.

During the parametric tests, the frequency was varied from 1 to 5 Hz, and the amplitude from 0.25 to 2 in. The drag was measured as a function of bag cap acceleration and amplitude for different frequencies. The data are presented in figures 8 and 9. A careful study of these data indicate that for low-sulfur coal fly ash the most effective operation in terms of drag versus acceleration is achieved at a shake frequency of 3 Hz and a shake amplitude of 2.0 in. This results in a bag cap acceleration of about 1.4 g.

Shake Duration

The effectiveness of shake/deflate cleaning depends on the length of the shaking period or the number of shakes imparted to the bag. A test was performed to evaluate the duration of shaking. Average drag was measured as the number of shakes was varied from 90 down to 9 shakes. At a frequency of 3 Hz, this corresponded to shake times of 30, 20, 15, 10, 5, and 3 s. The data are presented in figure 10. Below 5 s (~15 shakes) cleaning was ineffective in removing the dustcake and the drag degraded very badly. On the other hand, for shake times greater than about 20 s (60 shakes), there was not much additional drag reduction for the increased stress imparted to the bag.

Deflation Air-to-Cloth Ratio and Duration

Tests at the FFPP have demonstrated that unless the bags are slightly relaxed or deflated before shaking, very little dustcake is removed. To measure the amount of deflation volume needed for good cleaning action, drag was monitored as the deflation volume was varied from over 600 down to 0 ft^3. (The volume of an FFPP compartment is 4000 ft^3. The volume of all the bags in the compartment is 660 ft^3.) A deflation air-to-cloth ratio of 1 acfm/ft^2 was used during the test. The data are presented in figure 11. At least 100 ft^3, one-sixth of the total volume occupied by the bags, was required to achieve significant improvement in drag. Above this value, there was little additional reduction in drag until a fairly large volume was achieved. Overdeflation, which can accompany large deflation volumes, is not recommended since it can result in bag failures.

FILTERING AIR-TO-CLOTH RATIO

Parametric studies were conducted to develop relationships between filtering air-to-cloth ratio and tubesheet pressure drop for each bag cleaning method. These tests were performed for fixed filtering times of 60, 120, and 180 minutes. Filtering air-to-cloth values from 1.0 to 3.5 acfm/ft^2 were used. The resulting data are shown in figure 12. The cleaning sequences for the different cleaning methods were generally set up using the optimum conditions determined from the individual evaluations of the cleaning parameters. These results were described above. The figure shows that shake/deflate cleaning was able to maintain the lowest tubesheet pressure drop, followed by reverse-gas cleaning with sonic assistance and reverse-gas cleaning. While the values for reverse-gas cleaning are acceptable, the long-term expectation for this cleaning method would be for unstable operation and higher values of pressure drop.

These data have been used to calculate the relative economic benefits of operating at different filtering air-to-cloth values (1). These data are shown in figure 13. Levelized costs decrease as the filtering air-to-cloth ratio (A/C) increases up to 3.0 acfm/ft^2. This is higher than the A/C at which most utility baghouses operate. Also, the minimum in the curves occurs at a lower filtering air-to-cloth ratio for reverse-gas cleaning compared to the other two cleaning methods. Several points should be made concerning this graph. When capital costs and operating and maintenance costs are evaluated the result indicates that higher filtering air-to-cloth ratios than currently used at full-scale baghouses would result in cost savings. Allowing a "cushion" for operating upsets would indicate the selection of filtering air-to-cloth ratios for full-scale operation less than those corresponding to minimum cost as shown in figure 13. These would be 2.0 acfm/ft^2 for reverse-gas cleaning and 2.5 acfm/ft^2 for the other two cleaning methods.

FILTERING TIME

Using the previously selected filtering air-to-cloth values and using other operating data developed during the parametric tests, a computer model for baghouse pressure drop versus flow rate was used to evaluate average tubesheet pressure drop versus filtering time. These data are shown in figure 14. A minimum occurs in these data because as the filtering time decreases, the amount of time when compartments are out of service increases, thereby increasing the effective air-to-cloth ratio through the remainder of the baghouse. Several points should be made relative to this graph. First, the opacity will increase as filtering time decreases. This is because the majority of emissions occur immediately after cleaning. Second, the bag life will decrease as the filtering time decreases because of the accumulated additional stress on the bags. Third, these data then would suggest acceptable filtering times between 30 and 120 minutes; less than 30 minutes would generally not be possible at a full-scale baghouse and greater than 120 minutes would result in higher average pressure drops.

FABRICS

Historically, utility baghouses have used fiberglass bags with a 3 x 1 twill weave, a 10% Teflon B finish, and a warp-out construction with 75% of the exposed surface texturization facing the filtering (inside) of the bag. No data existed that proved that this was the best type of fabric for filtering coal fly ash. Since it was thought that the most likely variable strongly affecting the filtering capability of the fabric was exposed texturization, a study was undertaken to evaluate this parameter. A four-compartment, 2.5 MW(e), fabric filter pilot plant was used. Each compartment had essentially identical bags except for the texturization, which was 0, 25, 75, and 100%. Reverse-gas cleaning was employed and the filtering air-to-cloth ratio was about 1.6 acfm/ft^2. The baghouse was operated for 7000 h. The data are summarized in figure 15. While the drag was similar for each texturization, substantial differences were experienced in emissions and bag weight. These data suggest that fabrics with 25% exposed texturization can be used with certain coal fly ashes with operating results that are as good or better than for warp-out fabrics.

RESULTS OF DEMONSTRATION TEST

At the conclusion of the parametric testing, optimum operating parameters were selected for each cleaning method and the FFPP was configured for a "demonstration" test period. One compartment was set up for reverse-gas cleaning and one compartment was set up for shake/deflate cleaning. The other two compartments were used for reverse-gas cleaning with sonic assistance. In one of these two compartments sonic assistance was used every time the bags were cleaned. In the other compartment sonic assistance was used during every third bag cleaning cycle. The filtering time for all compartments was set at 90 minutes. The filtering air-to-cloth ratio in the reverse-gas-cleaned compartment was 2.0 acfm/ft^2. In the other three compartments it was 2.5 acfm/ft^2. A 3 x 1 twill weave fiberglass fabric with a 10%

Teflon B finish in the warp-in (~25% exposed surface texturization) orientation was used in all compartments. The cleaning parameters are summarized in table 1.

The FFPP was operated for 5000 h during this test. At regular intervals measurements were made of the average tubesheet pressure drop (drag), the residual dustcake weight (bag weights), and the emissions (outlet mass concentration). The performance data after 5000 hours of operation are presented in figure 16. Since the operating parameters were kept constant during the demonstration phase, the results of these tests are most representative of commercial operation.

The data for the reverse-gas-cleaned compartment shown in figure 16 compare well with the results from the fabric texturizations study presented in figure 15. At that time, for the 25% exposed surface texturization compartment after 7000 hours of operation, the drag was 2.7 in. H_2O/(ft/min), the residual dustcake areal density was 0.6 lb/ft^2, and the emission rate was 4×10^{-6} lb/10^6 Btu. At the end of the demonstration test the drag in the reverse-gas-cleaned compartment was 3.0 in. H_2O/(ft/min), the residual dustcake areal density was 0.6 lb/ft^2, and the emission rate was 2.5×10^{-5} lb/10^6 Btu. The comparability of these data from these two different long-term tests in separate baghouses demonstrates that stable operation had been achieved.

DRAG

The average drag in the reverse-gas-cleaned compartment was 3.0 in. H_2O/(ft/min). For the other three cleaning methods the average drag was 33% lower, 2.0 in. H_2O/(ft/min).

RESIDUAL DUSTCAKE AREAL DENSITY

The average residual dustcake areal density in the reverse-gas-cleaned compartment was 0.6 lb/ft^2. The reverse-gas-cleaned compartments with sonic assistance averaged 0.3 lb/ft^2, 50% lower than the reverse-gas-cleaned compartment. The average residual dustcake areal density in the shake/deflate-cleaned compartment was 0.25 lb/ft^2, 58% lower than the reverse-gas-cleaned compartment.

EMISSIONS

Emissions from all compartments were very low. For the reverse-gas-cleaned compartment the emission rate was 2.5×10^{-5} lb/10^6 Btu. The next lowest compartment was the one cleaned by reverse-gas with intermittent sonic assistance. The value was 1.5×10^{-4} lb/10^6 Btu. When sonic assistance was used with every bag cleaning the emission rate increased to 4.3×10^{-4} lb/10^6 Btu. The highest emissions were from the shake/deflate-cleaned compartment, which had emissions of 1.5×10^{-3} lb/10^6 Btu. All of these emission rates are well below NSPS regulations (3×10^{-2} lb/10^6 Btu).

CONCLUSIONS

This study of fabric filter operation has conclusively shown that, for collection of low-sulfur-coal fly ash from pulverized coal boilers, shake/deflate or sonic-assisted reverse-gas cleaning are preferable over traditional reverse-gas cleaning for utility baghouses. The use of warp-in (25% exposed surface texturization) constructed bags also aids in the maintenance of low residual dustcake areal densities on the filter bags.

Although these tests were not able to address the issue of bag life, there was no evidence of bag deterioration during the course of the tests, and during the numerous test periods over the eight years of operation of the Fabric Filter Pilot Plant, there have been no occurrences of bag failure other than damage due to handling. The longest period bags were in service was three years.

These pilot plant data can be used with full-scale data to estimate pressure drop versus filtering air-to-cloth ratio (A/C) for full-scale baghouses with seasoned dustcakes. This can be accomplished using the following simplified equations:

Tubesheet Pressure Drop = 3(A/C) ± 20% (For reverse-gas cleaning)

Tubesheet Pressure Drop = 2(A/C) ± 20% (For reverse-gas cleaning with sonic assistance or shake/deflate cleaning)

where A/C is expressed in actual cubic feet per minute per square foot and pressure drop is expressed in inches of water. Typically, at full-scale utility baghouses system or flange-to-flange pressure drop is measured instead of tubesheet pressure drop. These values are usually 1 to 2 in. H_2O higher than tubesheet pressure drops due to the ductwork pressure loss.

The lower value obtained using these equations would be representative of units with several months to a year of service; the upper value would be representative of units with two years or more of service. For example, for a reverse-gas-cleaned baghouse, assuming an A/C of 2.0 acfm/ft^2, tubesheet pressure drop of approximately 5-7 in. H_2O would be expected over time. If an additional 2 in. H_2O is added for ductwork pressure drop, a system pressure drop of 7-9 in. H_2O would be expected. For baghouses cleaned by reverse-gas with sonic assistance or the shake/deflate method, the corresponding system pressure drop would be in the range of 5-7 in. H_2O. In actual operation, account must also be made for removing compartments from service for cleaning and the reverse-gas flow used to clean the out-of-service compartments. The effect of these latter factors will be manifested in brief, periodic increases in system pressure drop, and the degree of increase will depend upon the number of compartments and the cleaning frequency.

It can be seen that in terms of system pressure drop (plant energy consumption) reverse-gas cleaning with sonic assistance or shake/deflate cleaning would be preferred. Since one inch of pressure drop corresponds to about $50,000 per year higher energy costs for a 500-MW plant, the potential savings during a normal 30-yr plant life can be substantial ($3,000,000). Currently, guidelines for predicting baghouse behavior in order to optimize their design and operation are being developed. These will include more detailed analysis of pressure drop, dustcake weight, fabric parameters, and the influence of the coal chemistry.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

ACKNOWLEDGMENTS

The research described in this paper was funded under EPRI Contract No. RP1129-8. The authors gratefully acknowledge the support of the Public Service Company of Colorado and the operators of the Arapahoe Test Facility, Raymond Kaiser Engineers. We wish to thank the Southern Research Institute scientists without whose efforts this work would not have been successfully completed, B. D. Hitchcock, G. E. Kenniston, R. H. Slye, and M. A. Smith.

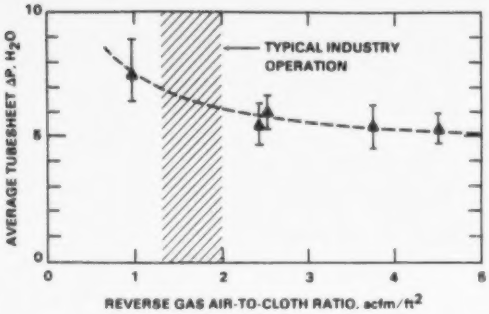

Figure 1. Relationship between tubesheet pressure drop and reverse-gas air-to-cloth ratio developed at the Fabric Filter Pilot Plant during parametric evaluation of reverse-gas cleaning.

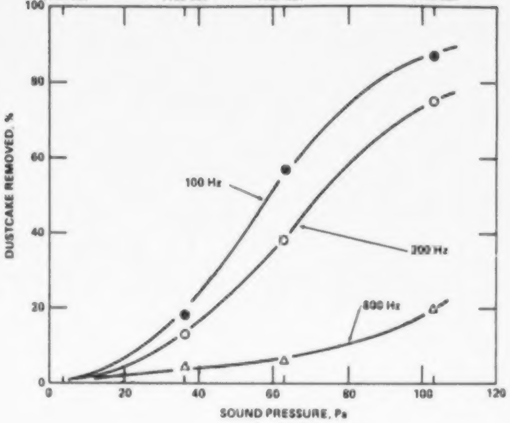

Figure 2. Summary of laboratory sonic cleaning experiments using fabric/dustcake samples from the Fabric Filter Pilot Plant. Dustcake removed after a 30-s exposure time versus sound pressure is shown for three frequencies.

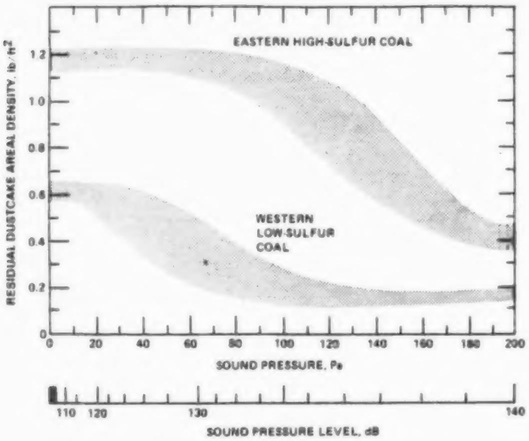

Figure 3. Effect of the local sound pressure in a baghouse compartment in reducing the residual dustcake from high- and low-sulfur coal. The power-weighted mean frequency was 211 Hz for all tests represented on this graph.

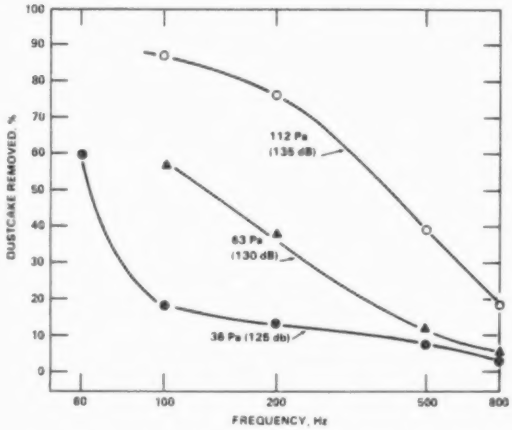

Figure 4. Summary of laboratory sonic cleaning experiments using fabric/dustcake samples from the Fabric Filter Pilot Plant. Dustcake removed after a 30-s exposure time versus frequency is shown for three sound pressures.

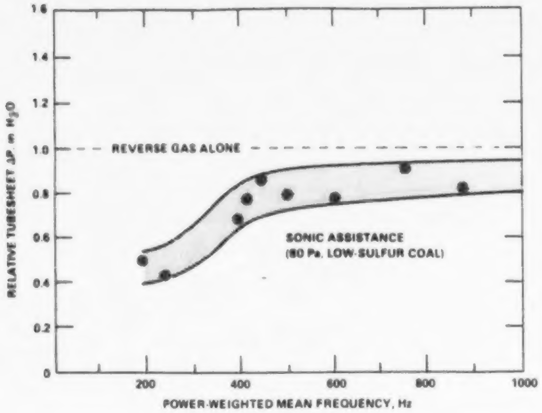

Figure 5. Effect of varying the power-weighted mean frequency of a horn on the reduction in pressure drop across the tubesheet of a fabric filter. All tests were made at an average sound pressure of 80 Pa.

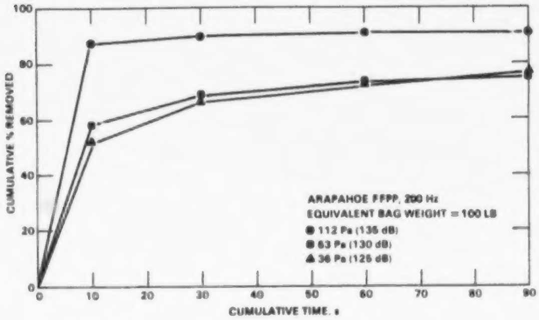

Figure 6. Relationship between cumulative exposure time and cumulative percentage dustcake removed during laboratory sonic cleaning tests. Low-sulfur fabric/dustcake samples from the Fabric Filter Pilot Plant were used. Three sound pressures were evaluated.

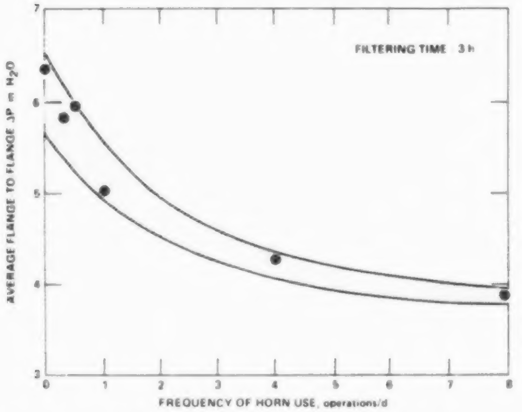

Figure 7. Effect of the operating schedule of horns on baghouse pressure drop. In all cases the horns were sounded during reverse-gas cleaning periods. Compartments were cleaned by reverse gas, with or without horns, every three hours.

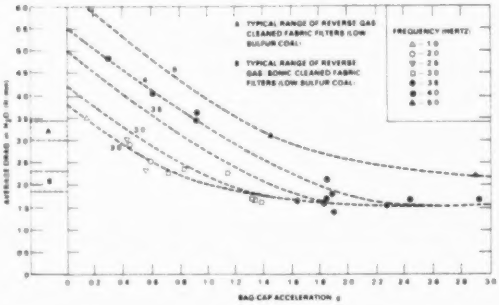

Figure 8. Average drag versus bag-cap acceleration for seven shaker frequencies. Data developed at the Fabric Filter Pilot Plant during parametric testing of shake/deflate cleaning.

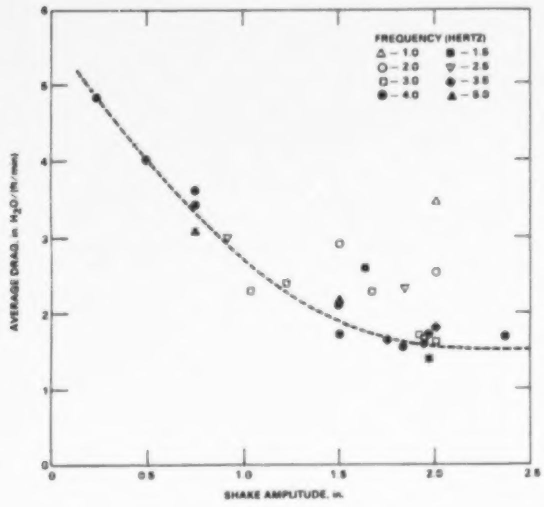

Figure 9. Average drag versus shake amplitude for eight shake frequencies. Data developed at the Fabric Filter Pilot Plant during parametric testing of shake/deflate cleaning.

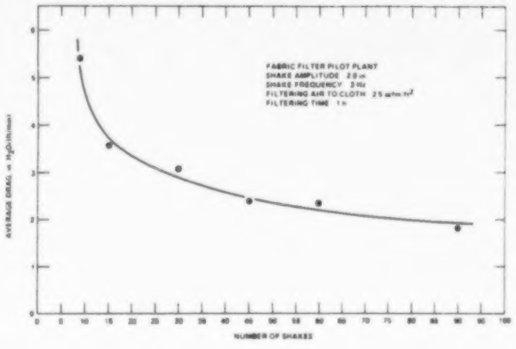

Figure 10. Average drag versus number of shakes at a shaker frequency of 3 Hz. Data developed at the Fabric Filter Pilot Plant during parametric testing of shake/deflate cleaning.

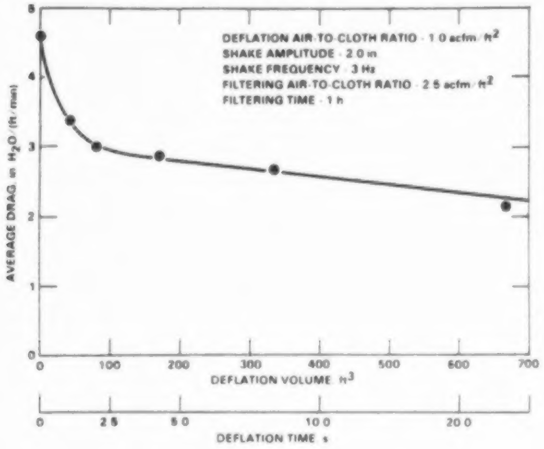

Figure 11. Average drag versus deflation volume. Data developed at the Fabric Filter Pilot Plant during parametric testing of shake/deflate cleaning.

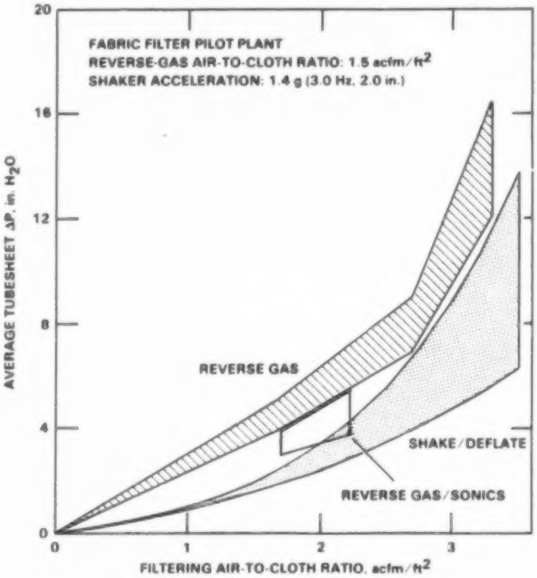

Figure 12. Average tubesheet pressure drop versus filtering air-to-cloth ratio for reverse-gas, reverse-gas with sonic assistance, and shake/deflate cleaning test periods at the Fabric Filter Pilot Plant.

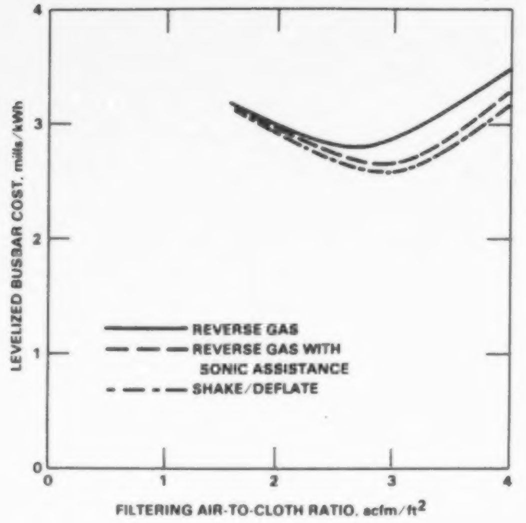

Figure 13. Levelized busbar cost versus filtering air-to-cloth ratio for three baghouse cleaning methods. Data for this figure were developed at the Fabric Filter Pilot Plant.

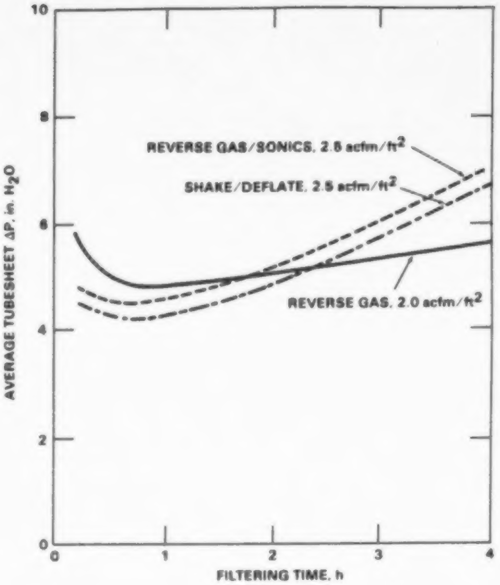

Figure 14. Relationship between average tubesheet pressure drop and filtering time for three baghouse cleaning methods. Data for this figure developed at the Fabric Filter Pilot Plant.

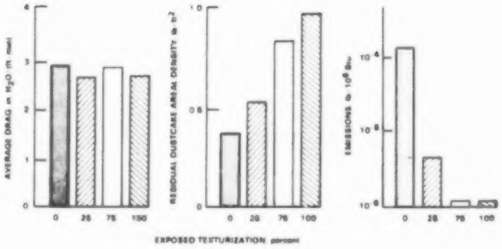

Figure 15. Effect of fabric texturization after 7000 h of operation during full-compartment testing at the Ecolaire baghouse.

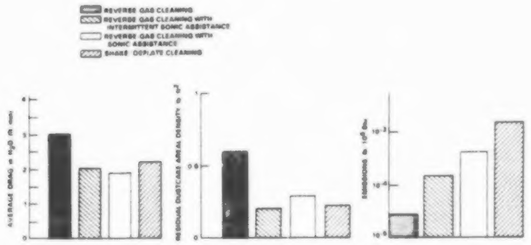

Figure 16. Final summary of three major baghouse performance indicators after 4500 h of operation of the Fabric Filter Pilot Plant during the Demonstration of Advanced Baghouse Operation.

TABLE 1. FFPP OPERATING PARAMETERS DURING DEMONSTRATION TEST

	Compartment			
	A	B	C	D
Air-to-Cloth Ratio, acfm/ft^2	2.5	2.0	2.5	2.5
Filtering Time, min	90	90	90	90
Reverse-Gas A/C, acfm/ft^2	1.5	1.5	1.5	1.0
Reverse-Gas Duration, s	30	30	30	5
Shake Amplitude, in.	-	-	-	2
Shake Frequency, Hz	-	-	-	3
Shake Duration, s	-	-	-	20
Fabric Texturization, %	25	25	25	25
Sonic Assistance	Yes	No	Every Third Cleaning	No
Average Measured Sound Pressure, Pa	97.6	-	96.8	-
Horn Duration, s	30	-	30	-
Cleaning Method	Reverse-Gas with Sonic Assistance	Reverse-Gas	Reverse-Gas with Intermittent Sonic Assistance	Shake/Deflate

FABRIC FILTER OPTIONS FOR HIGH-SULFUR COAL

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ABSTRACT

Design and operating options for fabric filters located at utilities burning high-sulfur coal have been investigated during research conducted from 1982 to 1987 at the EPRI 10-MW High-Sulfur Coal Fabric Filter Pilot Plant. This study revealed that residual dustcake weight, rather than pressure drop, was the most difficult parameter to control. Condensed sulfuric acid was found to be the primary cause of the high cohesivity of the ash which was, in turn, responsible for the heavy dustcakes. Minimizing acid dew point excursions reduced the rate of dustcake accumulation. The use of fabrics with low surface texturizations such as 25% and the use of energetic cleaning methods such as sonic-assisted reverse-gas cleaning and shake/deflate cleaning improved baghouse performance in terms of lower drags and residual dustcake weights. Both of the former cleaning methods were successfully demonstrated at an air-to-cloth ratio of 2.5 acfm/ft^2 with excellent particle collection characteristics. Based on these results, recommendations are presented for cleaning method, operating temperature, air-to-cloth ratio, and fabric texturization.

INTRODUCTION

As baghouses have gained wider acceptance as particulate control devices for utility boilers burning pulverized coal, it has become apparent that the performance of these baghouses depends upon the sulfur content of the coal. Although the Electric Power Research Institute (EPRI) had initiated a study of fabric filter performance at the low-sulfur coal fabric filter pilot plant (the FFPP described in the preceding paper), this site was unsuitable for investigation of the peculiarities of high-sulfur coal baghouse operation. EPRI therefore installed a pilot-scale baghouse at a power plant burning high-sulfur coal to increase the scope of its fabric research program because a significant portion of the utility industry uses high-sulfur fuels.

Research at this new pilot plant uses a multi-compartment baghouse with full-scale bags. The effort has produced a comprehensive examination of the three cleaning methods in general use at utility baghouses in terms of three indicators of baghouse performance: emissions, residual dustcake weight, and drag (the tubesheet pressure drop divided by the filtering air-to-cloth ratio). Testing began with reverse-gas cleaning because 90% of utility baghouses at the time the study began were cleaned by reverse gas. Later tests investigated sonic-assisted reverse-gas and shake/deflate cleaning. Initial testing of each cleaning method identified characteristics of baghouse performance associated with the cleaning method that required closer scrutiny. During these follow-on tests, exposed surface texturization (The exposed surface texturization (EST) is the percentage of texturized yarns exposed on the filtering side of the fabric.) was investigated and operation at an increased filtering air-to-cloth ratio (A/C) was demonstrated. Table 1 summarizes the research conducted since July 1982.

This paper summarizes baghouse performance observed at the high-sulfur pilot plant test facility. The following sections describe the attributes of the pilot plant essential to the research program and present an overview of the tasks performed in the first series of tests. Baghouse performance with recommended cleaning methods, exposed surface texturizations of the fabric of the bags, and operating parameters is described. Finally, conclusions and recommendations indicated by the results of the research program are summarized.

FACILITY DESCRIPTION

The research described here was performed at the EPRI 10-MW(e) High-Sulfur Coal Fabric Filter Pilot Plant (HSFP) hosted and cofunded by the Southern Electric System at the Scholz Electric Generating Station of Gulf Power Company in Sneads, Florida. This pilot-scale baghouse consists of five compartments in a parallel configuration with common inlet and outlet manifolds. The baghouse is pilot scale only in the sense that each compartment is limited to 42 bags. Full-sized bags and hardware are used to yield performance typical of commercial systems. The baghouse filters a slipstream of flue gas from a 40-MW pulverized-coal boiler burning eastern, bituminous coals containing 2 to 3% sulfur. To simplify data interpretation, the filtering air-to-cloth ratio is held constant. Reverse-gas (RG), sonic-assisted reverse-gas (RG/S), or shake/deflate (S/D) cleaning can be implemented. Pertinent baghouse specifications are listed in table 2.

DISCUSSION OF THE EXPERIMENT

Testing at the HSFP originally included tests to investigate a broad range of issues that were believed to be important for high-sulfur applications, including corrosion, acid attack of the bags, bag blinding, high pressure drop and others. However, as the tests proceeded, the program was focused on baghouse performance in terms of emissions, residual dustcake weight, and filter drag because these parameters proved to be the most important factors affecting the viability of fabric filters in high-sulfur coal applications.

Testing has been directed toward examining baghouse performance using three cleaning methods during steady state operation at a variety of boiler operating conditions. In addition, performance has been characterized during cyclic operation. Finally, the use of fabrics with differing percentages of texturization exposed to the dirty gas flow has been evaluated. Some results have been consistent throughout all of the work at the Scholz facility. Emissions have been low throughout. The fly ash is more cohesive than ash from typical low-sulfur coals, and forms a porous residual dustcake on the bags that has a relatively low drag. However, residual dustcake weight (areal density) has a tendency to increase in the absence of high-energy cleaning.

In fact, residual dustcake increased to more than 1.0 lb/ft^2 (equivalent to 100 lb net weight on the bags) in a few months with reverse-gas cleaning when the boiler operated at low load. Residual dustcake weight also increased following each baghouse outage for all but the most optimum cleaning methods. Since the baghouse inlet temperature during low load operation was near the acid dew point, it was concluded that transitions of the baghouse temperature through the acid dew point allowed sulfuric acid to condense, producing a highly cohesive dustcake.

The propensity of sulfuric acid to increase the cohesivity of a dustcake was further demonstrated in the laboratory. A sample of ash was washed several times with water to remove any water soluble acid. This sample and a second unwashed sample were mixed with water to form a slurry and dried in an oven. After drying, the unwashed sample exhibited much greater cohesivity than the washed sample, as evidenced by the photographs in figure 1. This result emphasized the need to maintain the baghouse operating temperature above the acid dew point to avoid rapid increases in residual dustcake weight.

The simulation of cyclic boiler operation (taking the baghouse off line each night and allowing it to cool below the dew point temperature) was found to dramatically accelerate the accumulation of residual dustcake. Some of the accumulation could be removed by returning to continual on-line operation, but most of it remained. This property of the ash was used to advantage later in the test program to quickly build a "mature" dustcake (one more typical of high-sulfur coal baghouses) on new bags.

REVERSE-GAS AND SONIC-ASSISTED REVERSE-GAS CLEANING

As stated earlier, reverse-gas cleaning, the most commonly used cleaning method in utility baghouses serving pulverized-coal boilers, was tested first. Conservative air-to-cloth ratios typical of utility baghouse practice in 1982, (1.6 and 1.8 acfm/ft^2 net (with all compartments in service)) were employed during these tests. Bags with 75% EST were used for this initial testing, and the selection of cleaning parameters (such as reverse-gas A/C and duration) was based on the results of earlier studies conducted at the FFPP. The boiler operated at low loads throughout the test. Baghouse operation included periods of steady-state operation, no-purge-shutdown cyclic operation (the baghouse was not vented of flue gas when taken off-line), and operation following a simultaneous baghouse/boiler start-up. Throughout all of these tests, low drag (1.6 to 1.9 acfm/(ft/min)), low emissions (0.0002 lb/10^6 Btu), and heavy residual dustcake weights (often exceeding 1 lb/ft^2) were found to be typical characteristics of operation with reverse-gas cleaning. These results indicated that more energetic cleaning was required to reduce residual dustcake weight and that operation at higher filtering air-to-cloth ratios was feasible, since drag for the tenacious but porous dustcake was low.

Since heavy residual dustcakes persisted with reverse-gas cleaning, an evaluation of sonic-assisted reverse-gas cleaning followed. This method of cleaning was chosen because it had been demonstrated to be a more efficient cleaning method in a number of utility installations and is retrofittable to reverse-gas units. The initial test of RG/S at Scholz was conducted for an A/C of 1.8 acfm/ft^2, 75% EST fabric, and with boiler operation at low loads. During this test, drag remained between 1.4 and 1.7 in. H_2O/(ft/min), and residual dustcake weight was limited to less than 0.5 lb/ft^2.

Follow-on testing of RG and RG/S was conducted at high boiler load in conjunction with a fabric texturization study to evaluate the performance of these cleaning methods using bags with 25% exposed surface texturization and a reverse-gas A/C of 1.5 acfm/ft^2. Following 8130 h of operation, drag averaged 2.2 in. H_2O/(ft/min) in the RG/S compartment (A/C = 2.5 acfm/ft^2) and 2.5 in. H_2O/(ft/min) in the RG compartment (A/C = 2.1 acfm/ft^2). At the same time, residual dustcake weight was 0.45 lb/ft^2 in the RG/S compartment and 0.85 lb/ft^2 in the RG compartment. The significant advantage, especially in dustcake weight, of sonic-assisted reverse-gas cleaning identified it as an attractive option for high-sulfur coal applications.

A deterioration in horn performance from a compartment average sound pressure of 104 Pa at startup to 83 Pa at the conclusion of the evaluation of this cleaning method was noted. The drop in sound pressure is believed to be due to corrosion within the horn driver assembly, caused by acid condensation during the period of cyclic operation and during normal startup and shutdown activities. Appropriate installation in a full-scale baghouse should minimize corrosion of the horns. Even with the decrease in horn performance, RG/S was able to control drag and dustcake weights at desirable levels.

SHAKE/DEFLATE CLEANING

Extensive tests of shake/deflate cleaning were also conducted, and this cleaning method was shown to be a useful alternative for high-sulfur coal baghouses. Since results from previous parametric tests were not available for S/D, extensive testing was required to determine acceptable operating parameters. Over the two-year period of evaluation, with a filtering A/C of 2.5 acfm/ft^2 and bags with 75% EST, the parameters significantly affecting this cleaning method were examined in detail.

Initial work with shake/deflate cleaning revealed that deflation A/C and duration were important to the successful application of this cleaning method. Improper deflation was determined to be the probable mechanism causing the failure of 17% of the HSFP bag complement after less than one year of S/D operation. After replacing the bags, further evaluation of S/D determined minimums for deflation parameters that would maximize bag life and control both residual dustcake weight and drag. During 7440 h of operation with the proper deflation, the second complement of bags experienced no failures. Using these deflation settings, filtering time, filtering air-to-cloth ratio, bag-cap acceleration, and shake duration were studied. The parameters determined to be optimum are identified later in this paper. During the shake/deflate parametric tests, the baghouse experienced two serious upsets which increased drag or dustcake weight or both: a boiler tube leak and an unscheduled no-purge outage. The baghouse was able to recover to the previous levels of drag and dustcake weight by continuing normal operation with shake/deflate cleaning.

The study culminated in a period of boiler operation at high load when S/D was demonstrated at a filtering air-to-cloth ratio of 2.5 acfm/ft^2 with a drag of 2.0 in. H_2O/(ft/min), a residual dustcake weight of 0.36 lb/ft^2, and emissions of 0.002 lb/10^6 Btu.

EXPOSED SURFACE TEXTURIZATION STUDY

The degree of surface texturization of the fabric was shown to have a strong effect on dustcake removal during cleaning at tests performed by EPRI at the Arapahoe Test Facility. An evaluation of fabric texturization was subsequently performed at the HSFP. The purpose of this test was to compare average drag and residual dustcake weight for bags made of fabrics with 0, 25, and 75% EST mounted in adjacent RG/S compartments.

The evaluation led to a better understanding of the ways in which boiler operation affects baghouse performance. During the first 4800 h of operation, residual dustcake weight in all three of the RG/S compartments and drag in the two compartments with texturized bags increased very slowly because the boiler had adopted a predominantly high-load operating schedule. The higher baghouse temperatures (>300°F) associated with the high-load operation minimized excursions through the acid dew point, resulting in low residual dustcake weights. The drag in the compartment with non-texturized bags increased rapidly following baghouse outages, rising at times to values twice those for the texturized fabrics.

A period of cyclic baghouse operation was conducted, with the baghouse removed from service for 8 h each day, to build a mature dustcake quickly. Residual dustcake weight increased as much as 540% during the 657 h the baghouse was on line (16 h/day). Drag for the texturized fabrics changed little, but the average drag for the non-texturized fabric decreased by 30%, resulting in about the same drag for all three compartments. The dustcake weight measured at the end of the period of cyclic operation was more typical of a high-sulfur coal baghouse than the weights measured at the end of the first 4800 h of operation. A period of continuous operation followed, and drag and residual dustcake weight increased slowly, reaching equilibrium after about two months. At the conclusion of this post-cyclic period which lasted for 2680 h of operation, drag had increased significantly: 25, 40, and 50% for the bags with 0, 25, and 75% EST, respectively. In comparison, the residual dustcake weights increased very little: 6, 15, and 0% for the bags with 0, 25, and 75% EST, respectively. The absence of a direct correlation between increases in residual dustcake weight and drag during the first period of continuous operation and the final period of operation was a consequence of changes in dustcake porosity associated with the dew point excursions. Figure 2 compares the drags and residual dustcake weights associated with the three fabric texturizations at the end of each of the three operating periods.

The effect of acid dew point excursions on residual dustcake weight is further demonstrated in figure 3, where the residual dustcake weight is plotted against time and number of outages for the pre-cyclic and cyclic operation periods. A stronger correlation exists with number of outages (dew point excursions) than with time of operation.

Another interesting observation was made during the period of cyclic operation. The drag for the bags with 0% EST, which at times had been up to twice those for the other fabrics, fell until it was the lowest of the three. It is thought that the increased ash cohesivity also produced increased porosity. However, average drag in the remaining compartments remained essentially unchanged, even though significant increases in residual dustcake weight resulted from the cyclic operation. Further studies will attempt to uncover the reason for this behavior.

Although the fabric with 0% EST had the lowest drag and lightest dustcake at the end of the test, a concern about the possibility of diminished bag life with 0% EST fabric remains. Four failures occurred with bags made from this fabric which are believed to have been caused by internal abrasion in areas of very light dustcake. All of the failures developed in bags adjacent to the horn in the area of the bags within 2 ft of the bell mouth. The problem should not be overstated, since the failures occurred after more than a year of operation and none of the other bags in the compartment failed. Consequently, for boilers which experience cyclic operation or significant numbers of outages, fabrics with no texturization may be a viable economic alternative. Further studies of these fabrics are planned.

OPTIONS FOR HIGH-SULFUR COAL BAGHOUSE OPERATION

The EPRI high-sulfur coal fabric filter research program has led to the identification of several options for improved baghouse operation and performance in high-sulfur coal applications. These options include the selections of operating air-to-cloth ratio, type of fabric for bags, and cleaning method to be used in the baghouse. Various options and the anticipated baghouse performance associated with them are described in this section. The careful application of these options in a baghouse design or modification program should result in significantly improved performance.

Operating at temperatures below the acid dew point (experienced during low load operation and baghouse startups) has caused significant increases in dustcake weight. The possibility of making adjustments in boiler operation that do not affect power production economics but would be beneficial to baghouse operation by keeping the baghouse temperature above the acid dew point should be considered.

Low drag measured throughout testing at the HSFP has indicated that higher filtering A/C values are viable options for future baghouses. The data collected during the evaluations of RG and RG/S demonstrated that a high-sulfur coal baghouse cleaned by RG/S can operate successfully at an A/C of 2.5 acfm/ft^2.

Reverse-gas cleaning alone may not be the preferred choice for a high-sulfur coal baghouse. Although the drag was moderate in comparison with low-sulfur coal baghouses, concern remains for the high residual dustcake weight and its implications for reduced bag life. In initial testing, the residual dustcake weight exceeded 1 lb/ft^2 for bags with 75% EST fabric. During later tests with bags having an exposed surface texturization of 25%, dustcake weights were still 0.85 lb/ft^2. The heavier dustcakes and the higher drag (2.5 in. H_2O/(ft/min) compared to 2.2 and 2.0 in. H_2O/(ft/min) for RG/S and S/D, respectively), obtained with a fabric with better dustcake release properties (25% EST rather than 75% EST used in the first RG test), confirmed the conclusions of earlier research that reverse-gas cleaning is not desirable for future high-sulfur coal applications.

Sonic-assisted reverse-gas cleaning was shown to be a viable option for use in a high-sulfur coal baghouse. RG/S was demonstrated with 25% EST bags, a filtering A/C of 2.5 acfm/ft^2, and a reverse-gas A/C of 1.5 acfm/ft^2. The measured compartment average sound pressure of 104 Pa at startup decreased to 83 Pa by the end of the test. The test covered 8130 h of operation at mostly high boiler load, but included 657 h (41 days) of cyclic baghouse operation. When operated concurrently with a compartment cleaned by RG and having the same fabric, the RG/S compartment had 14% lower drag (2.2 in. H_2O/(ft/min)), 16% higher A/C, and 47% lower dustcake weights (0.45 lb/ft^2). This cleaning method can therefore be recommended for retrofitting existing reverse-gas baghouses as well as for future installations.

This research program has also shown that shake/deflate cleaning can control drag and residual dustcake weight to acceptable levels in a high-sulfur coal baghouse. S/D cleaning was demonstrated using bags with 75% EST and a filtering A/C of 2.5 acfm/ft^2. The maximum bag-cap acceleration for HSFP shakers, 1.7 g with a bag-cap shaking amplitude of 1.5 in. at a frequency of 3.7 Hz, was needed for successful operation. Proper deflation (the minimum required for adequate pressure drop reduction) was necessary for successful cleaning and suitable bag life. With S/D, the baghouse recovered from two serious upsets: a boiler tube leak and an unscheduled no-purge outage. After 7440 h of operation with the boiler operating at high loads, average drag was 2.0 in. H_2O/(ft/min) and residual dustcake weight was 0.36 lb/ft^2. Figure 4 summarizes baghouse performance with the three cleaning methods.

Bags made with less texturization exposed to the dirty gas flow are a clear choice for use in high-sulfur coal baghouses because they give lower residual dustcake weight and drag. After 8130 h of operation, including 41 days of cyclic operation to accelerate the accumulation of residual dustcake, the compartments with 0 and 25% EST had lower drags and residual dustcakes about one-half of those on the bags with 75% EST. Emissions were somewhat higher with the lower texturizations but remained more than an order of magnitude lower than the 1978 new source performance standard for stationary sources (0.03 lb/10^6 Btu). Figure 5 shows drag, residual dustcake weight, and emissions for each of the three fabric texturizations included in this investigation.

CONCLUSIONS AND RECOMMENDATIONS

Research at the High-Sulfur Coal Fabric Filter Pilot Plant has encompassed three cleaning methods and three fabric options. Particulate collection efficiency varied for the differing fabrics and cleaning methods but was excellent for all of them. High levels of cleaning energy were needed to control the accumulation of the cohesive dustcake formed by high-sulfur coal combustion. Reverse-gas cleaning produced low-to-moderate drags but produced high residual dustcake weights. Sonic-assisted reverse-gas cleaning and shake/deflate cleaning produced low-to-moderate drags and successfully controlled residual dustcake weights. For the different fabric texturizations, both residual dustcake weight and drag varied directly with exposed surface texturization; dustcake weight was more greatly affected by increased texturization. Major test results are summarized in table 3.

These and other conclusions drawn from the EPRI high-sulfur coal fabric filter research program allow the following recommendations to be formulated:

- Industry should reconsider the use of reverse-gas cleaning for future high-sulfur coal baghouses because residual dustcake weight is not controlled adequately by this cleaning method.

- For existing reverse-gas-cleaned baghouses, the installation of sonic horns and the use of fabrics with improved dustcake release characteristics can be expected to improve performance.

- Dustcake weight has increased dramatically following baghouse shut-downs leading to the conclusion that excursions through the acid and/or water dew point temperature should be minimized.

- For new facilities, baghouses using sonic-assisted reverse-gas or shake/deflate cleaning provide viable particulate control options.

- Bags with 25% EST are the best choice for a high-sulfur coal baghouse at this time. This texturization offers the best combination of drag, residual dustcake weight, emissions, and projected bag life. However, for some applications, such as continuous or frequent cyclic operation, bags with no surface texturization may be preferable.

- Higher filtering air-to-cloth ratios should be considered for future RG/S baghouses. Successful operation at an A/C of 2.5 acfm/ft^2 has been demonstrated at pilot scale for a high-sulfur coal baghouse cleaned by RG/S.

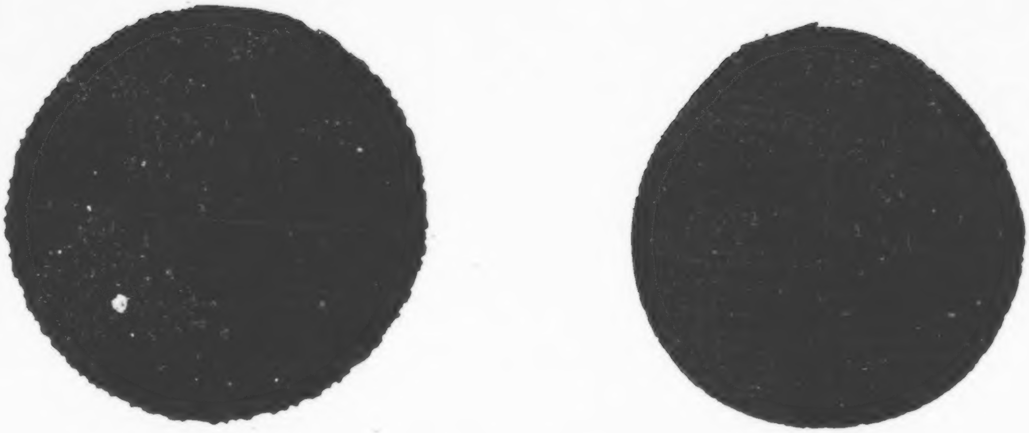
As a result of this and other EPRI research, design and operating parameters for high-sulfur coal fabric filters have been established, and are included in Fabric Filters for the Electric Utility Industry, a series of EPRI publications.

Future work at the HSFP will include: (1) a 2-MW(e) scale program combining heat pipe heat exchanger research with an extensive program of fabric research and development (This program includes periods of operation with the heat exchanger upstream of the baghouse to allow the development of fabrics for low-temperature operation.), (2) a program to study the effects of injecting low concentrations of various materials that will react with the SO_3 present in the flue gas (Both ammonia and calcium based sorbents will be studied.), and (3) a concurrent program of accelerated bag life evaluation using frequent shake/deflate cleaning to simulate years of wear in a few months of operation.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

ACKNOWLEDGMENTS

The work described in this paper was performed for the Electric Power Research Institute within the scope of project number RP 1867-4 under the direction and management of Mr. W. V. Piulle. The authors also acknowledge the contributions of Southern Company Services, Gulf Power Company, and the management at Scholz Electric Generating Station, all of which were vital to the success of this work.



WASHED **UNWASHED**

Figure 1. Photographs of ash cakes from washed and unwashed bag ash. The ashes were dried overnight at 110°C and the resulting ash cakes were slightly agitated.

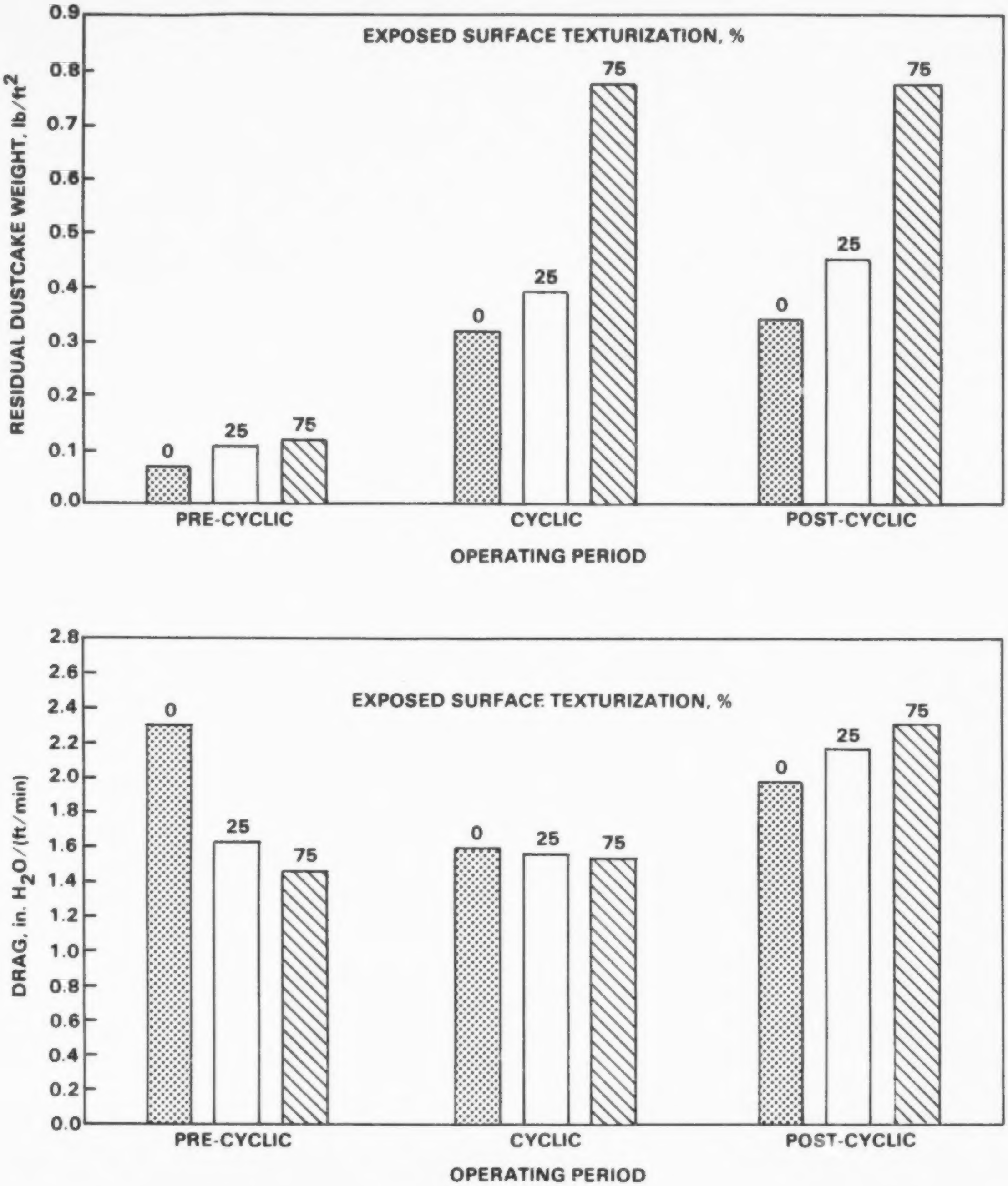
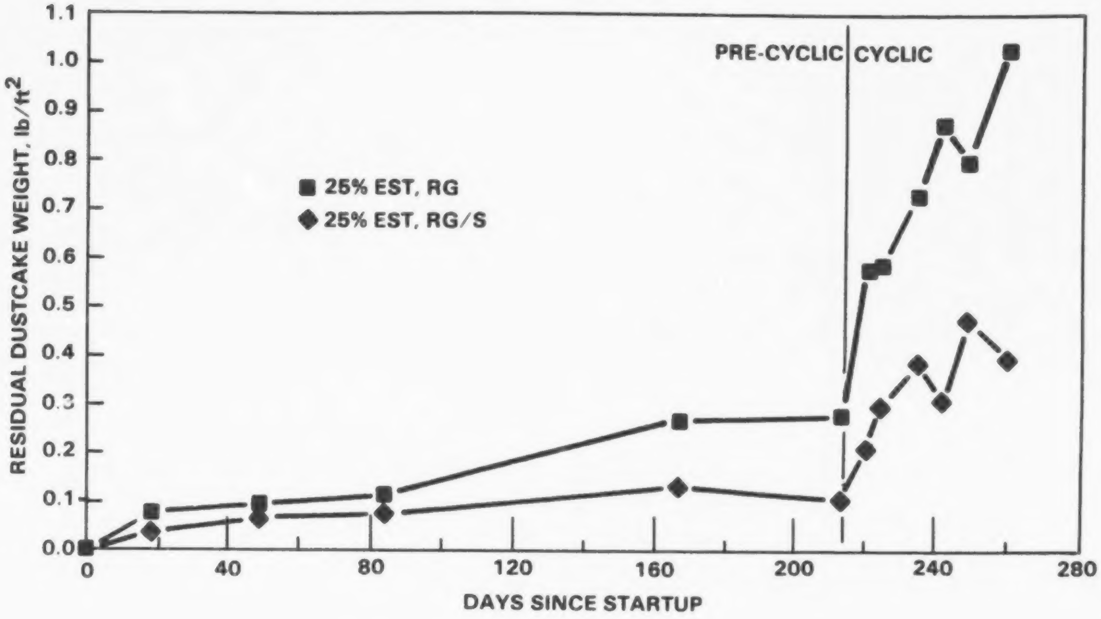


Figure 2. Residual dustcake weights and drags for the three sonic-assisted reverse-gas-cleaned compartments during the three major operating periods.



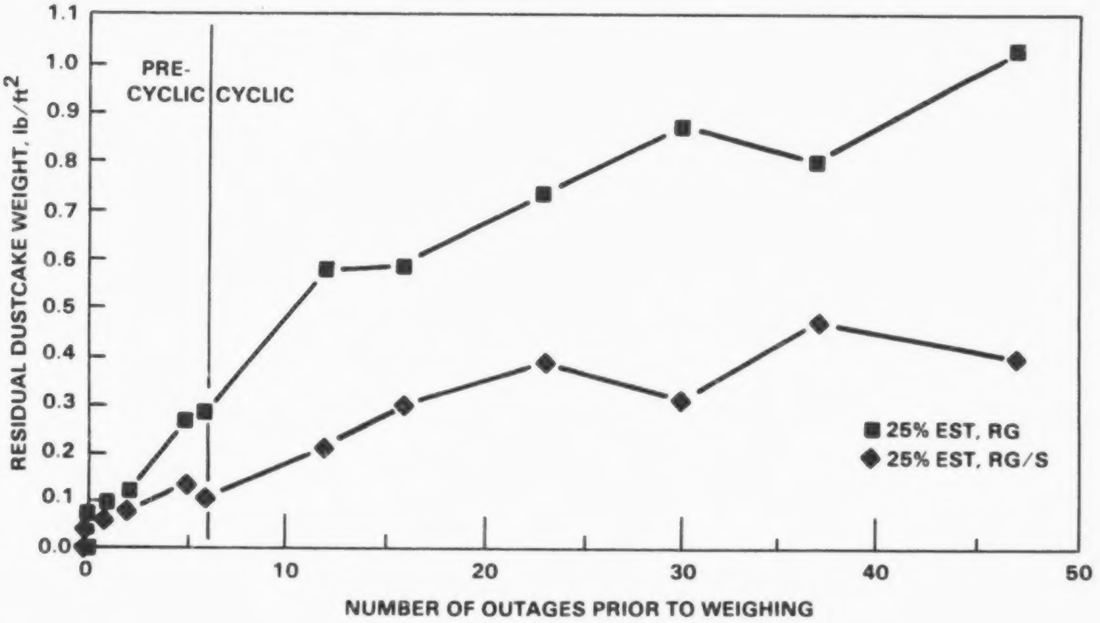


Figure 3. Residual dustcake weight as a function of time and of outages (dew point excursions).

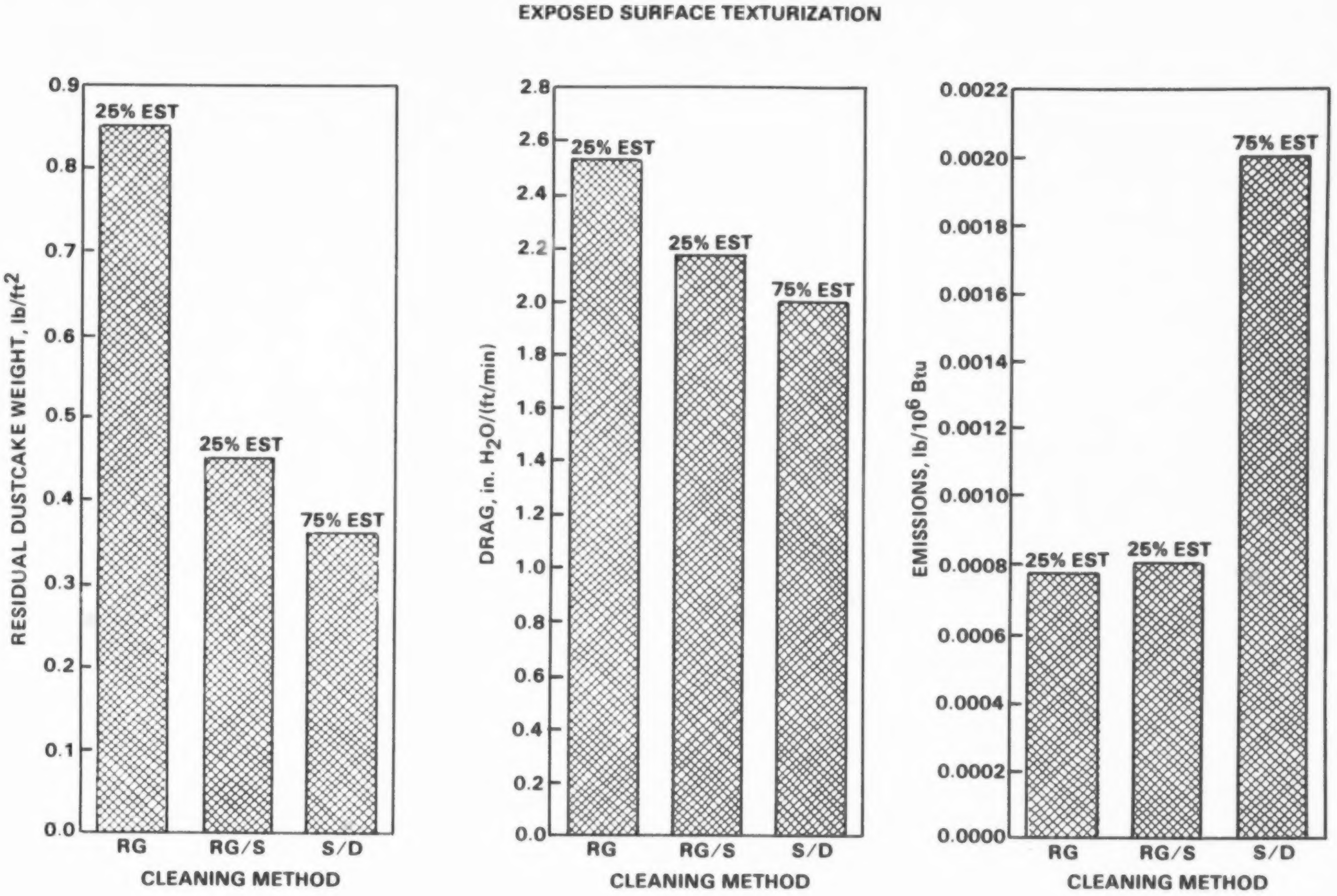


Figure 4. Residual dustcake weight, drag, and emissions for the three cleaning methods.

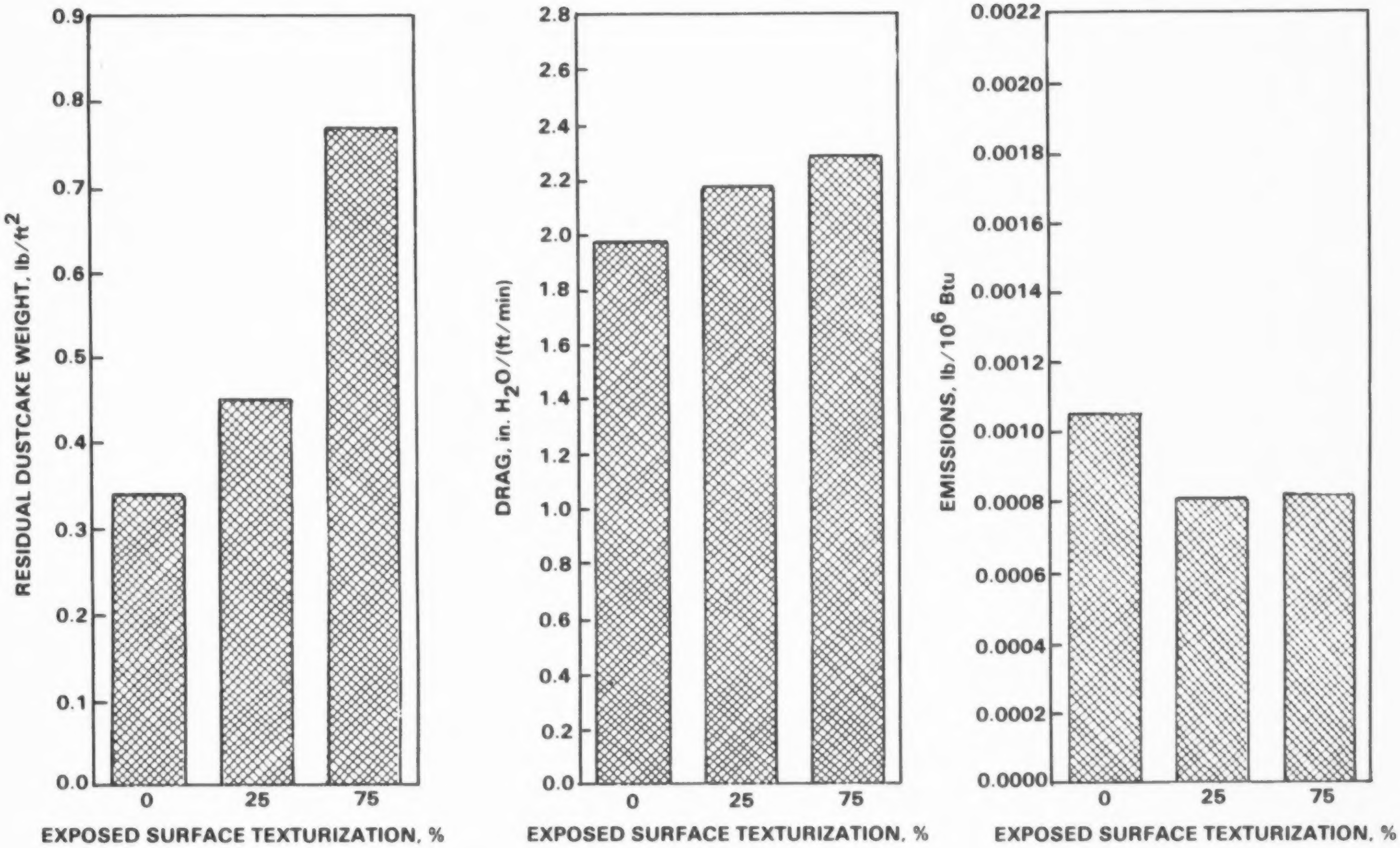


Figure 5. Residual dustcake weight, drag, and emissions for the three fabric texturizations.

TABLE 1. CHRONOLOGICAL SUMMARY OF HSFP RESEARCH

	Cleaning method	Air-to-cloth ratio ($acfm/ft^2$)	Year
Preliminary tests	Reverse gas	1.6 1.8	1982 1983
	Sonic-assisted reverse-gas	1.8	1983-84
	Shake/deflate	2.5	1984-85
Follow-on testing	Shake/deflate	2.5	1985-86
	Reverse-gas and sonic-assisted reverse-gas with fabric texturization	2.1 2.5	1986-87

TABLE 2. HSFP BAGHOUSE SPECIFICATIONS

Manufacturer	GEESI
Number of compartments	Five
Bags	Forty-two per compartment, 12 in. diameter by 35 ft long, fiberglass, variable fabric weaves and orientations.
Cleaning method	Reverse-gas, sonic-assisted reverse-gas, or shake/deflate.
Baghouse instrumentation	Flange-to-flange pressure drop, inlet temperature, outlet temperature, inlet duct static pressure, outlet duct static pressure, inlet mass concentration, reverse-gas flow rate, reverse-gas duct static pressure, reverse-gas temperature, outlet opacity, outlet mass concentration.
Compartment instrumentation	Tubesheet pressure drop, flow rate, inlet temperature, outlet temperature
Manual testing	Inlet and outlet mass concentrations, inlet and outlet particle size distribution, inlet and outlet SO_2/SO_3 concentrations, fly ash and coal analyses.

TABLE 3. MAJOR TEST RESULTS

	Air-to-cloth ratio (acfm/ft^2)	Particulate Emission	Drag	Dustcake Weights
Reverse gas	1.6 1.8 2.1	Low	Low-to-moderate (compared to low-sulfur units)	High
Sonic-assisted reverse-gas	1.8 2.5	Low	Reduced by up to 14% from reverse gas	Reduced by up to 50% from reverse gas
Shake/deflate	2.5	Low	Reduced by up to 21% from reverse gas	Reduced by up to 60% from reverse gas

	Fabric Exposed Surface Texturization		
	0%	25%	75%
Drag	Low	Low	Low/moderate
Dustcake Weight	Low	Low/moderate	High
Particulate Emissions	Low	Low	Low

EXPERIENCE WITH HIGH TEMPERATURE FILTER MEDIA IN ENVELOPE TYPE FILTERS IN FLUE GAS CLEANING

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ABSTRACT

In the past 5 years considerable field experience has been obtained with envelope type fabric filters in the cleaning of flue gases of combustion operations.

Fuels range from wood, coal, peat to lignite and waste, used to raise hot water and steam.

Flue gas temperatures varying from 100 deg. C. - 250 deg. C. (212 deg. F. - 482 deg. F.) permitted the use of a wide variety of available high temperature filter media.

Economic results and fabric durability varied widely depending on operating factors, boiler operation and maintenance practices.

Lately further experience has been gained in the reduction of SOx and HCl through the injection of special large specific area hydrated lime in the flue gases from the combustion of coal and municipal solid waste.

SOx reductions in the order of 65% and HCl reductions in the order of 95% and more have been obtained with a true dry scrubbing method.

As a result of the rapidly growing interest in The Netherlands in late 1982 to generate heat and create hot water and steam through the combustion of solid fuels, DCE became involved in the collection of solid contaminants from flue gases. Now over 30 installations have been built and operated in greenhouses and industry on combustion plants using coal, lignite (brown coal), wood, straw, RDF and industrial waste as a fuel.

Stokers which were used included screw feeders, chaingrate, stepped grate, pulverized fuel and fluid bed combustors. Whereas hot water and steam at various pressures are being generated, flue gas temperatures vary between wide limits which called for the use of different high temperature filter fabrics. Considerable experience has been gained with all high temperature filter fabrics available in the market place.

New users of coal fired boilers have to ensure that the following limits will not be exceeded before permits to install such boilers will be granted by the local authorities.

The Ministry of the Environment now imposes the following emission limits for coal fired combustors of grate and fluid bed type.

Pulverised fuel less than 300 MW capacity :

Dust : 50 mg/Nm3 - 0.022 g/scft
SOx : 700 mg/Nm3 - 245 ppm
NOx : 500 mg/Nm3 - 245 ppm
dry gas 7% 02, 288 gr. K, 101.3 Kpa, 60% RH.

Pulverised fuel greater than 300 MW capacity :

Dust : 50 mg/Nm3 - 0.022 g/scft
SOx : 400 mg/Nm3 - 140 ppm
NOx : 400 mg/Nm3 - 195 ppm
dry gas 6% 02, 288 gr. K, 101.3 Kpa, 60% RH.

It is evident that those requirements for max. allowable dust emissions can not be met with dust collectors other than EP's or fabric collectors. EP's however were not a serious consideration for the smaller size industrial boilers - up to approx 30 MW.

Although the fabric collector would provide the low emission rates without too much of a problem, little knowledge was available with regard to the behavior of different filter fabrics available on the world market in this application.

Although in the USA the application of fabric collectors in different shapes and forms on boiler plants in municipal power stations had progressed considerably through the years, it appeared this approach could not easily be transplanted into European situations.

Collectors were of different construction and using cleaning mechanisms not commonly used in Europe. In addition municipal power stations boilers operate under more or less constant conditions over a prolonged period of time. This was not the case with most industrial boilers for steam or hot water, where loads can vary considerably depending on the application.

It appeared that the automatic reverse jet fabric collector would be the most suitable type of separator available to tackle the problem. The operation of a reverse jet fabric dust collector is based on a simple technique. Dust laden gases are forced to pass through a textile material where the solids are separated and remain on the outer surface of the fabric. The clean gases pass through the filter material into the clean gas manifold. As the dust cake on the filter fabric increases, the resistance to gas flow increases, necessitating reconditioning of the media from time to time.
By blowing compressed air in short bursts counter to the gas flow into the filter element, the dust cake is broken up and falls into the dust hopper.

Diagram 1. shows the construction of the Dalamatic(tm) reverse jet envelope type collector. The construction is rather unique to the American market and its features need some explanation.

The collector is of modular construction with horizontally inserted filter envelopes 10 or 15 sq ft filter area each.
Each module contains 10 filter elements - 100 or 150 sq ft in total, side by side. Module dimensions are approx 3'4" wide by 1'8" high and which can be stacked up to 8 high and arranged in an unlimited number wide. The construction is very compact and usually lends itself to indoor installation.
Filter element removal and inspection is from the front, clean air side. Dirty gas inlet is at the top, clean gas outlet as well. The downward flow pattern avoids redeposition of the dust and supports dust settlement in the hopper. Compressed air controls are in the lower portion of the filter construction serving vertical jet tubes each handling one vertical row of envelopes at anyone time. The filter utilizes felted fabrics.

The most important component of a fabric dust collector is obviously the textile filtering material. Filter fabrics are sensitive products made from fine natural or synthetic fibers. In the selection of the appropriate fabric serious consideration has to be given to the chemical and physical properties of the dust as well as the composition and temperature of the carrier gases.

Dust properties can directly influence the collection results. They can also affect the service life of the filter material.
Different types of coal and boilers produce fly-ash of varying composition and behavior. The abrasive properties of the fly-ash should also be considered and have their influence on collector configuration, design inlet velocities and filter element arrangement.
The top inlet of the DCE Dalamatic collector, creating a down flow in the collector housing, turned out to be of prime importance to avoid flotation and re-entrainment of the non-agglomerating fly-ash. This allowed operating at fairly high filtration velocities without affecting the operating pressure drop adversely.

In all combustion operations the sulphur content of the fuel is of great importance to predict acid dew points in the flue gas stream. Acid condensation was expected to cause clogging and chemical attack of the filter media as well as corrosion in general. The acid dew point, which depends on the sulpur content of the coal and the water content of the flue gases is extremely difficult to calculate in advance. Certain quantities of sulphuric acid often are neutralized by the alkaline contents of the fly-ash. As a consequence the acid dew point may vary between 50 deg. C. (122 deg.F.) for a low sulphur, high alkaline coal and 130 deg. C. (266 deg.F.) or more for a high sulphur coal. We experienced far less clogging due to acid condensation than we had expected.

Initially quite a few boilers were installed in greenhouses to create hot water for heating purposes. Natural gas prices were at their peak in that period and firing coal turned out to be very economic. These installations, as a result of the quickly varying boiler loads, presented major operating problems :

- drop in flue gas temperatures leading to acid condensation
- incomplete combustion of coal leading to media blinding through tarry products causing high pressure drops
- spark entrainment at low loads - fire risks
- poor maintenance

Industrial boilers suffer from far less of these problems as loads are more constant and technically qualified maintenance personnel are usually available.

Diagram 2. shows a typcial equipment arrangement for these type of installations with spark arrestor cyclone, automatic bypass system and fabric dust collector. Cyclone and filter housing are usually insulated.

At this moment 36 DCE Dalamatic™ filter plants have been installed for the cleaning of flue gases in solid fuel fired boilers, hospital incinerators, crematoria and several other projects will appear shortly, particularly in incineration.

Table I shows a variety of boiler/feeder types that have been equipped.

Diagram 3 shows the relevant position of different fabrics available for high temperature filtration.

All filter fabrics - in felted form - shown in table II were utilized at one time or another.

The most common cause of media failure has been found to be chemical and hydrolytical attack.

The operating conditions of dust collectors and fabrics in greenhouses were far from ideal and would have shown much better media life under constant operating conditions as are typical for industrial boilers.
With a few exceptions all installations in the greenhouses were passing through acid dew points at least several times a day. Some industrial installations cool down to ambient temperature during week ends, not only causing acid condensation but moisture condensation as well.

Fabric operating experience can be summarized as listed in table III. Media life would be assumed to be acceptable if it showed at least 12 - 18 months reliable performance, with the exception of the more costly fibers such as Teflon or Tefaire which should show at least 24 - 36 month life.

Filtration velocities (AC ratios) were in the range from 3 - 8 fpm depending on filter media utilized and type of fuel being burned.

As mentioned before, frequent acid and occasional moisture condensation in the filter housing and on the fabric could not be avoided. Hot water boilers have flue gas outlet temperatures of 200 - 250 deg. F. industrial steam boilers equiped with economizers may be in the same range. These conditions caused glass felt to break down after 3 - 4 months.

Also Nomex (chemically treated) showed a shorter life than originally anticipated. Approximately 12 month was the longest life that could be achieved. High water vapor and acid content of the flue gases were responsible for the relatively short life. Much better life was obtained in installations operating under constant gas temperatures above the acid dew point or
where small quantities of hydrated lime were injected into the filter, specifically Nomex and PAC perform better under those circumstances.

Acceptable life was also obtained in installations burning lignite (brown coal) as a fuel, as a result of the relatively high alkaline content of the fuel and ashes. We used Nomex chemically treated in these cases.

Ryton has performed satisfactorily in some installations in spite of the limitations in temperature and oxygen content of the flue gases recommended by the fiber producer.
In a recent case however, Ryton-Ry failed completely within 6 month, probably because of high bromine content of the coal.

Teflon and PTFE fabrics behaved as expected, although some heat setting problems were encountered. The limitations in its use so far are largely caused by the cost of the fabric and the relatively low a/c ratio that should strictly be adhered to.

Polyacrylnitrile (Dralon T) showed good life initially, however, we have no encounter problems with failuresafter 6 - 9 month of operation, due to unclear reasons, which could be the use of higher sulphur coal, mild winter and/or a combination of those factors.

A little less than two years ago Du Pont introduced Tefaire felt media in the European market. One year before that time we agreed with Du Pont to run tests with Tefaire on a coal fired boiler installation in a greenhouse.
In these tests, over a period of 16 weeks, we increased the filtration velocity from 4.0 fpm to 10.0 fpm by reducing the number of filter envelopes in the filter housing.

The increase in Δp is plotted against the actual filtration velocity in diagram 4.

The test results were very encouraging and since the introduction of this new felt, 4 filter installations have been equipped The media enables the construction of small filter plants thus off-setting its relatively high cost in comparison with lower temperature media's.

Diagram 5 shows the relative filter cost flange to flange when equipped with different media.

Diagram 6 shows life expectancy of the different filter media that have been tested in actual field installations.

Recently the interest in the hot gas filtration market re-activated, this time however, flue gas cleaning projects concern combustors using Refuse Derived Fuel (RDF), municipal solid waste (MSW), hospital waste and crematoriums.
Some of these applications also require dry scrubbing with activated lime to eliminate not only particular contaminants but also noxious gases such as HCl, HF and SO2, but also heavy metals and polyaromatic components.

A few months ago, tests were completed for the filtration and dry scrubbing of flue gases from a MSW waste combustor using a new highly active lime hydrate with over 40 m2/g (28 sg.ft/gr) area. (standard hydrated lime is approx 50% of this)
Results are shown in diagram 7 and diagram 8.

As you will note very good reduction of HCl and SO2 emissions in particular could be achieved with a simple technique and at a stoichiometric ratio of 1.5. These systems require no recirculation of partly spent hydrated lime and operate completely dry.

In all these installations the envelope type filter performed excellently with :

- filter fabric pressure drop : 2 - 8" WG
- dust inlet into filter : 0.2 - 5 gr/cuft
- dust emissions : less than 0.02 gr/cuft

Recently DCE has introduced a new envelope type filter element - the Sintamatic - installed in the filter housings in the same way as described before. The Sintamatic element consists of a high grade porous composite with a Teflon based coating. It is a rigid self supporting element, offering extremely high filtration efficiency.

Future elements will certainly go in the direction of such elements, offering considerable advantages over the classical felts and weaves used in high efficiency dust filtration today.

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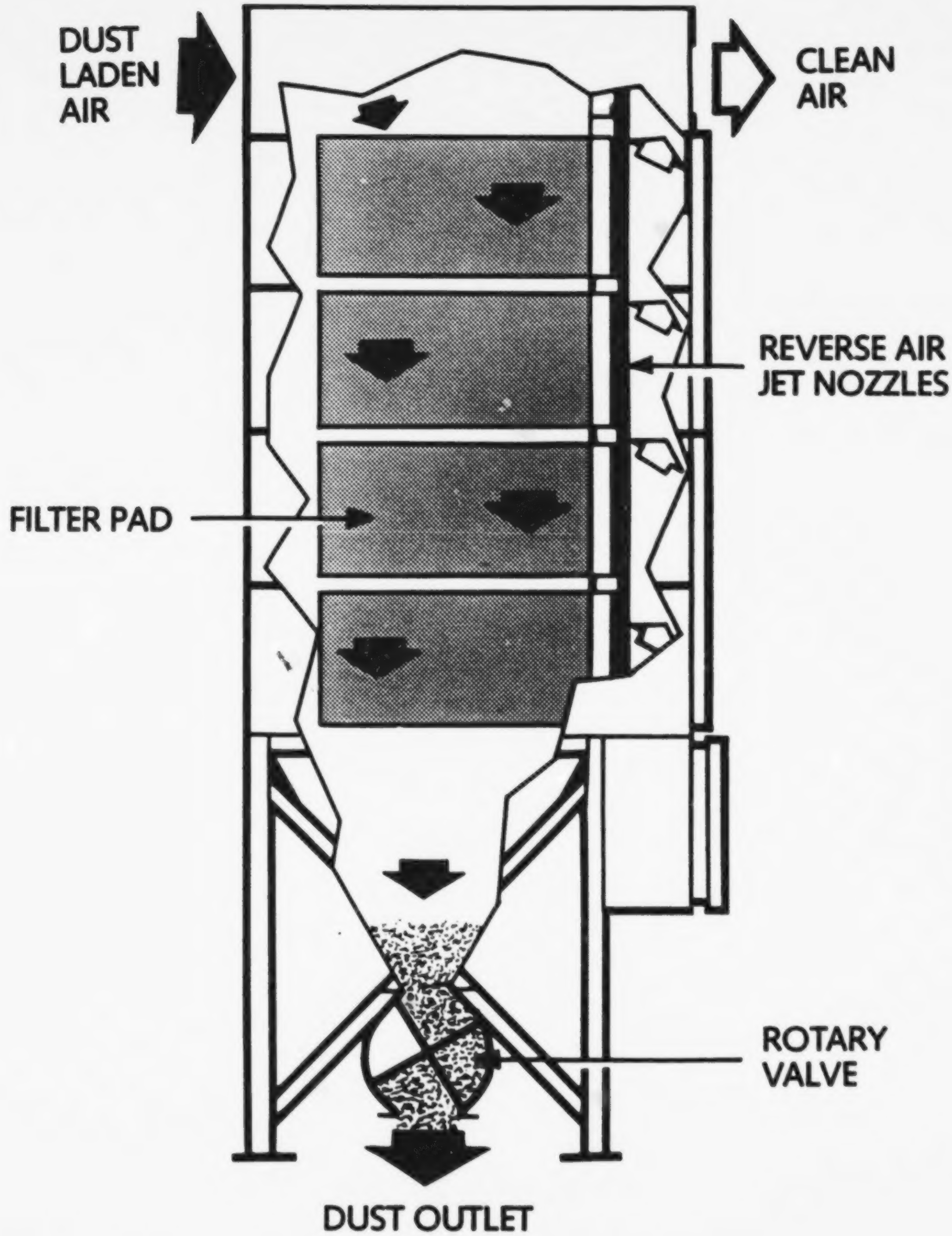


FIG. 1 IDEAL DESIGN SHOWING DUST ENTERING FILTER AT TOP

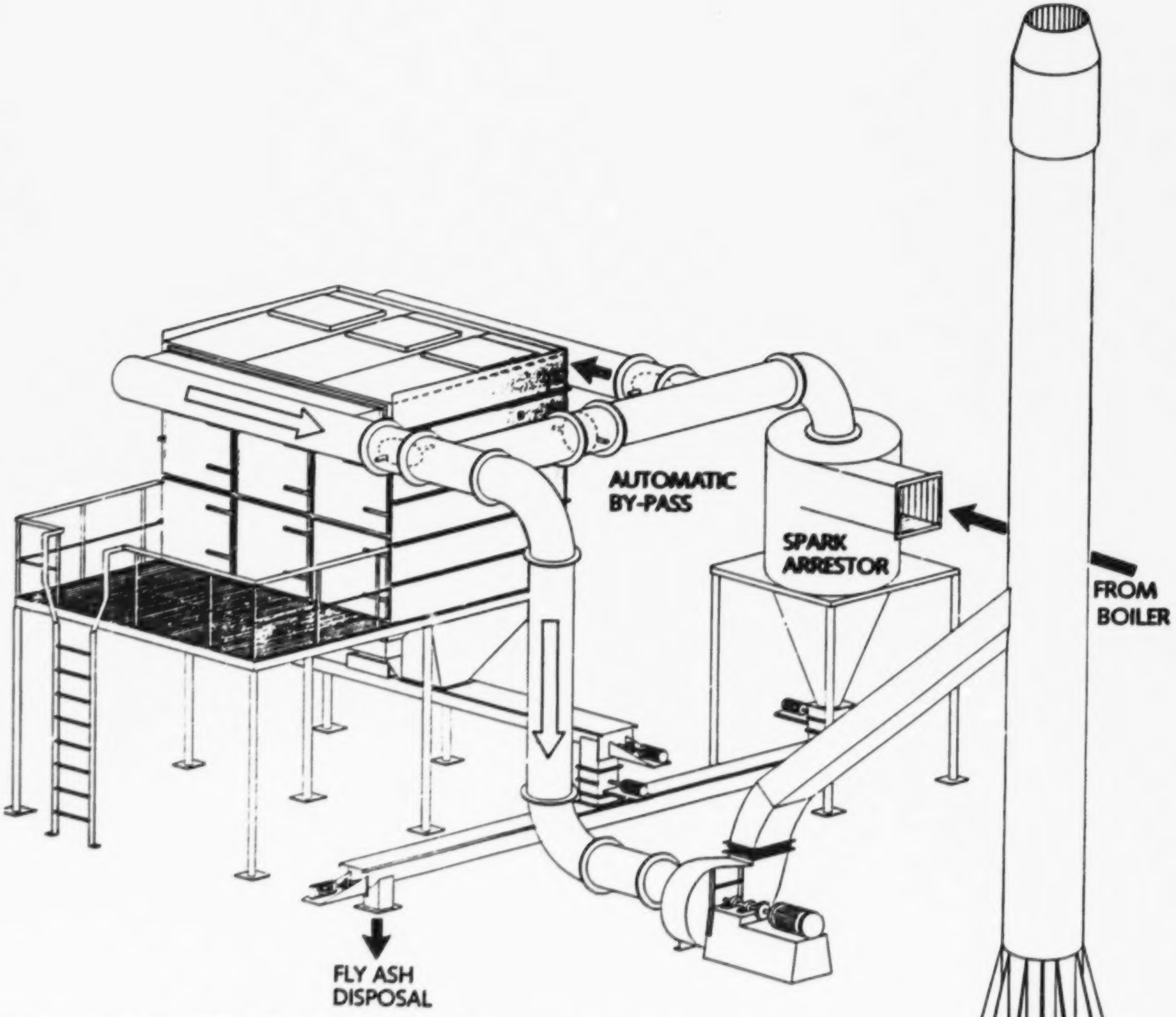


FIG. 2 DALAMATIC DUST COLLECTOR ON BOILER PLANT

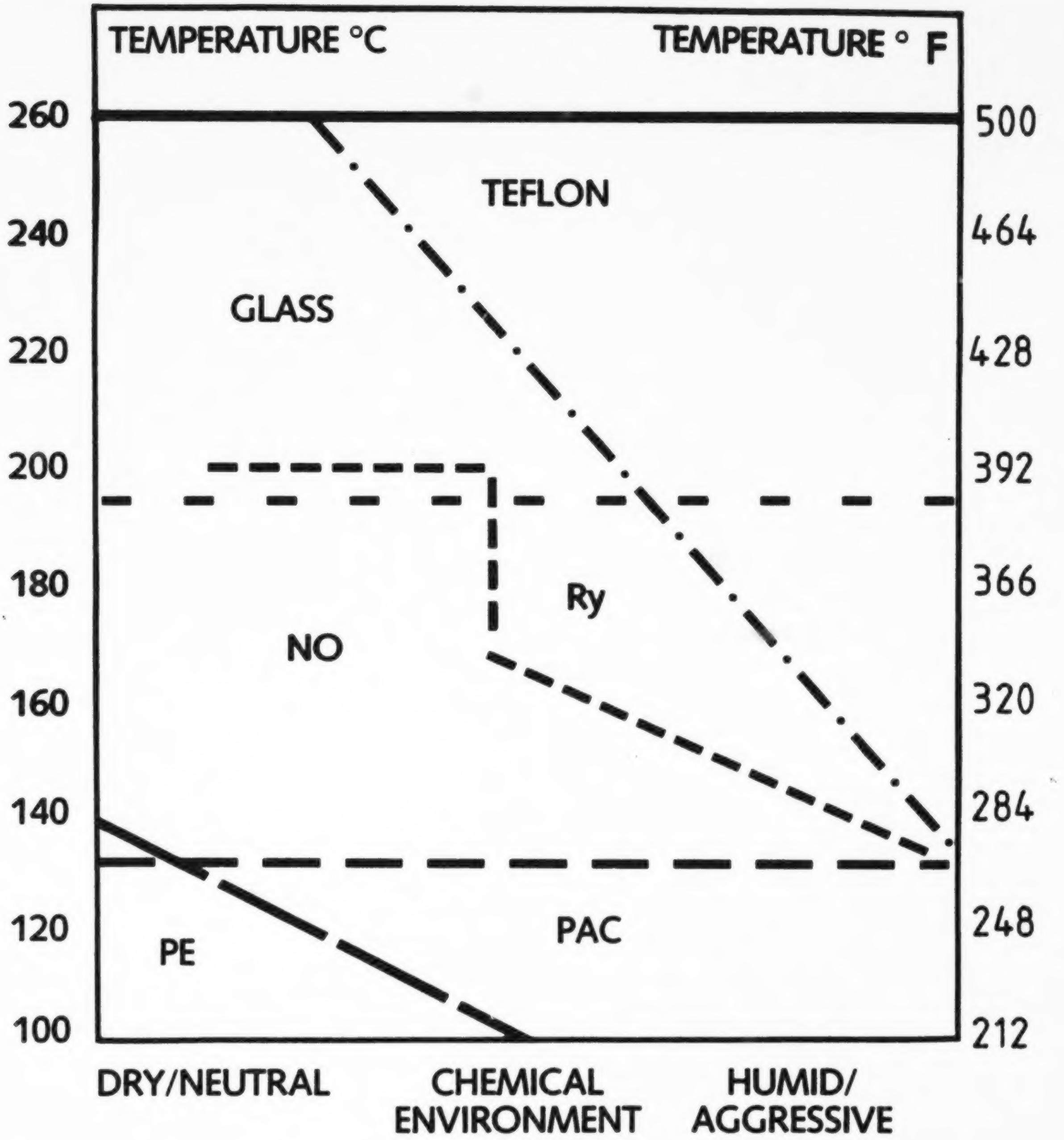
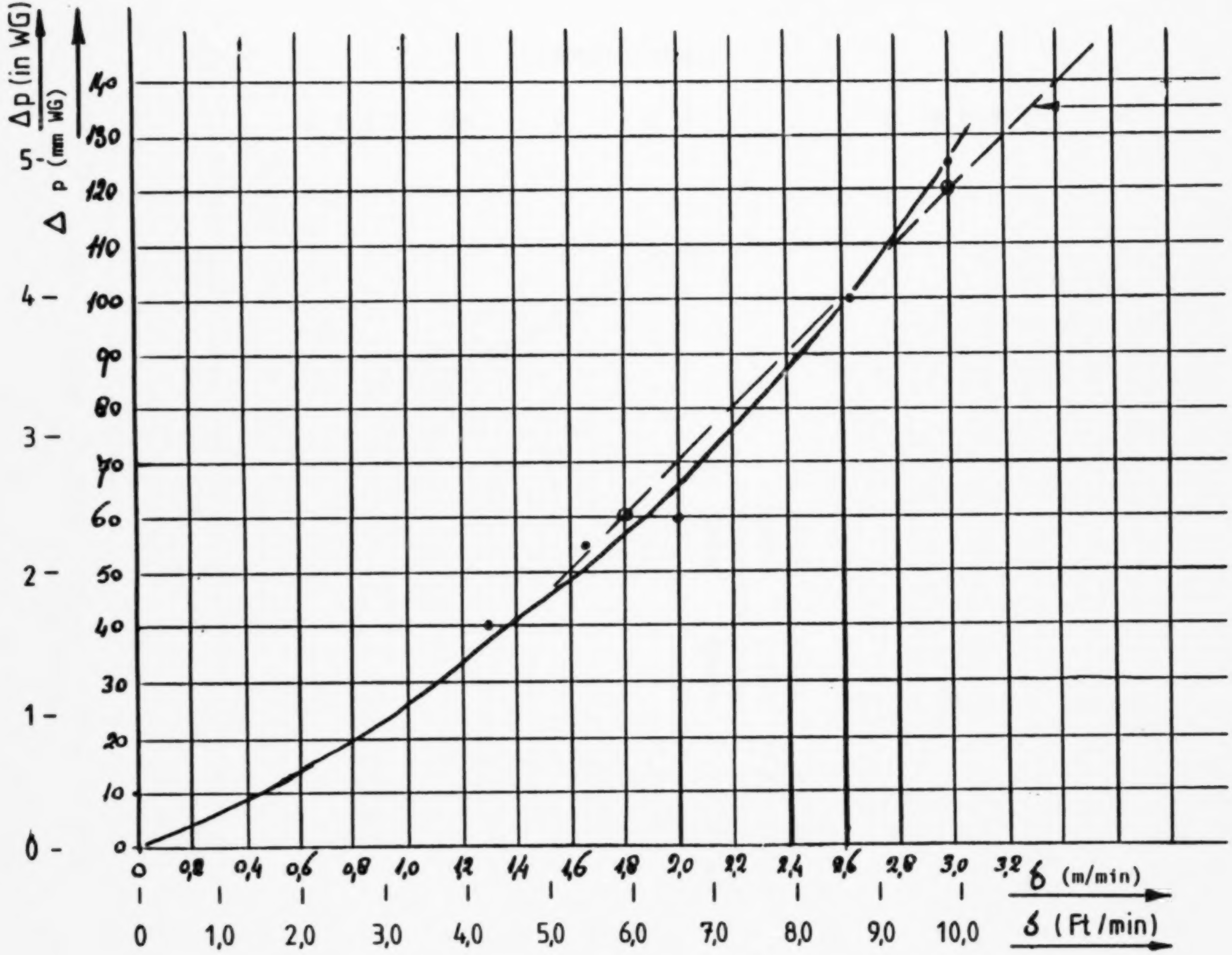


FIG 3. PRODUCT POSITIONING

SOURCE
DU PONT

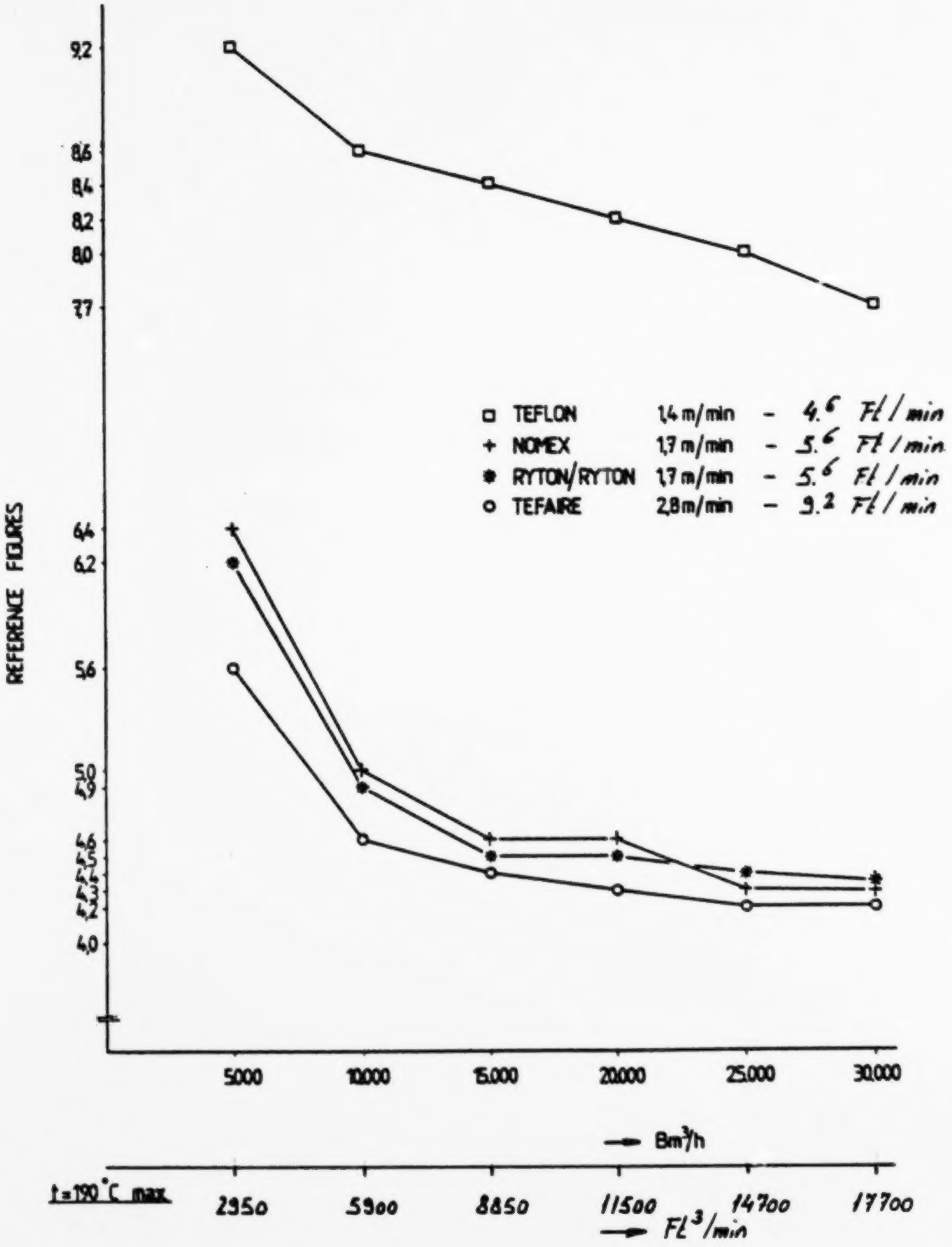


Publication Du Pont

Figure
4

TEFAIRE

Figure 5. Comparison DCE Dalamatic filter prices with different high temp. media's.



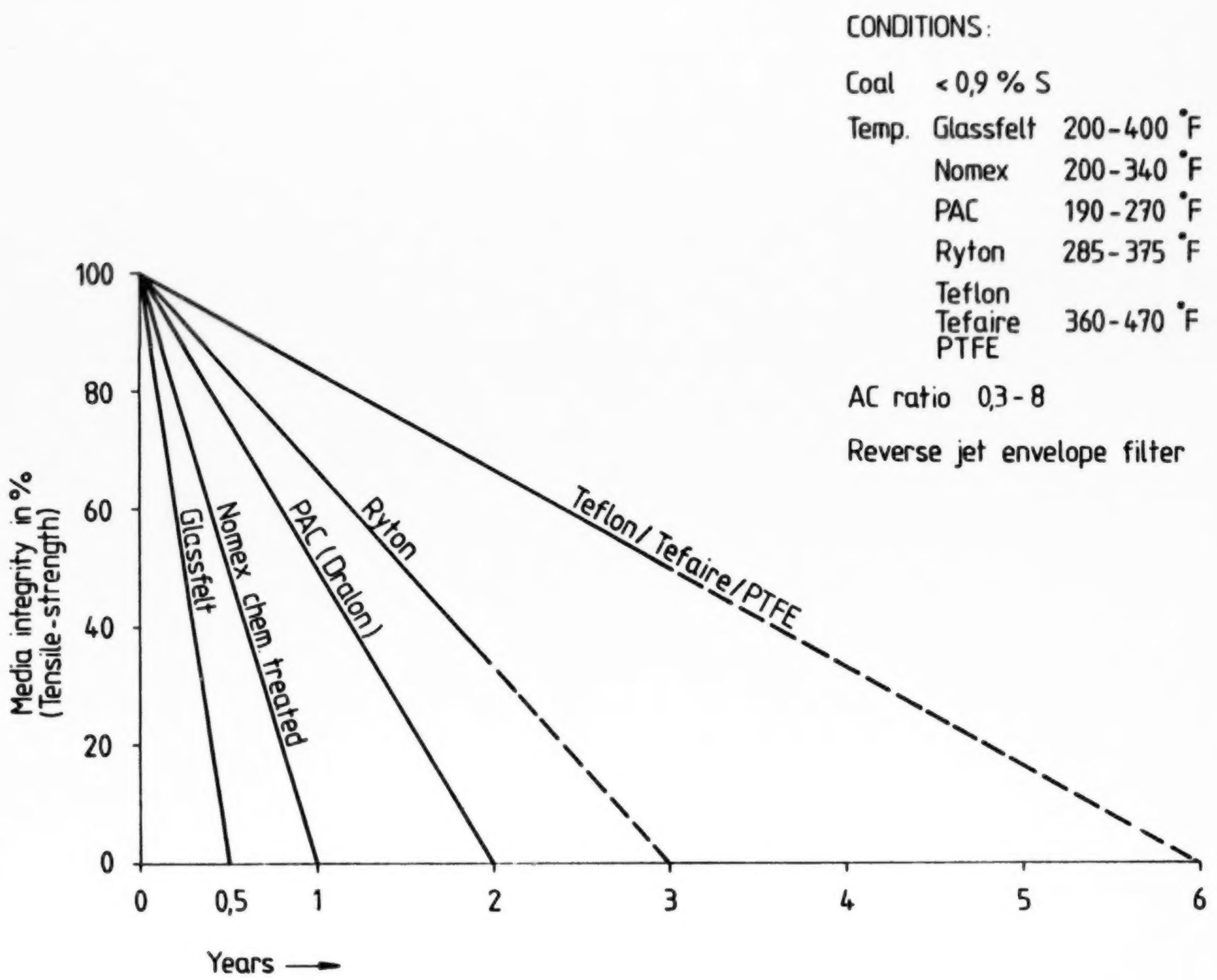


FIGURE 6. LIFE EXPECTANCY HIGH TEMP. FILTERMEDIAS IN FLUEGAS CLEANING

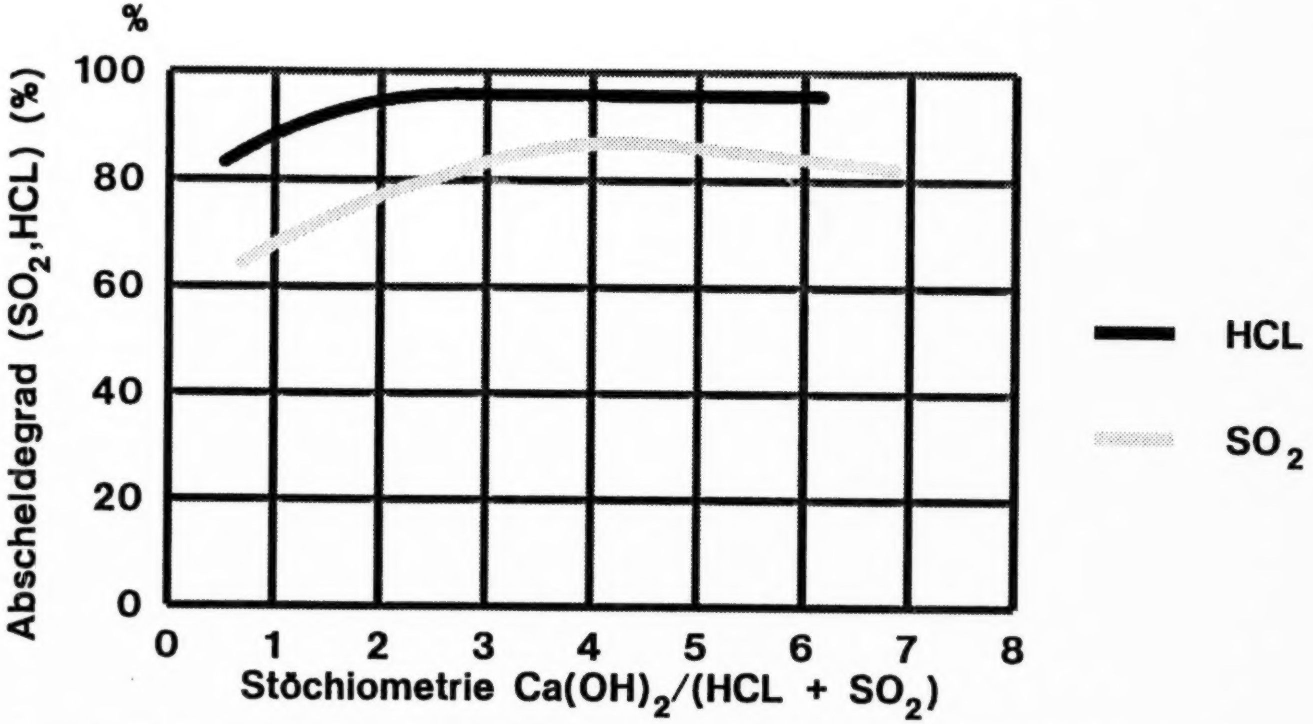


Figure 7 **Gemittelte (aus mehr als 300 Einzelmessungen) Abscheidegrade von HCL und SO_2 als Funktion des stöchiometrischen Faktors (bezogen auf HCL + SO_2)**

Wasserdampfgehalt der Abgase:	**ca. 14%**
Reaktionstemperatur:	**ca. 130°C**
Durchschnittlicher Rohgasgehalt:	**590 mg $HCL/m^3{}_n$** **280 mg $SO_2/m^3{}_n$**

DCE

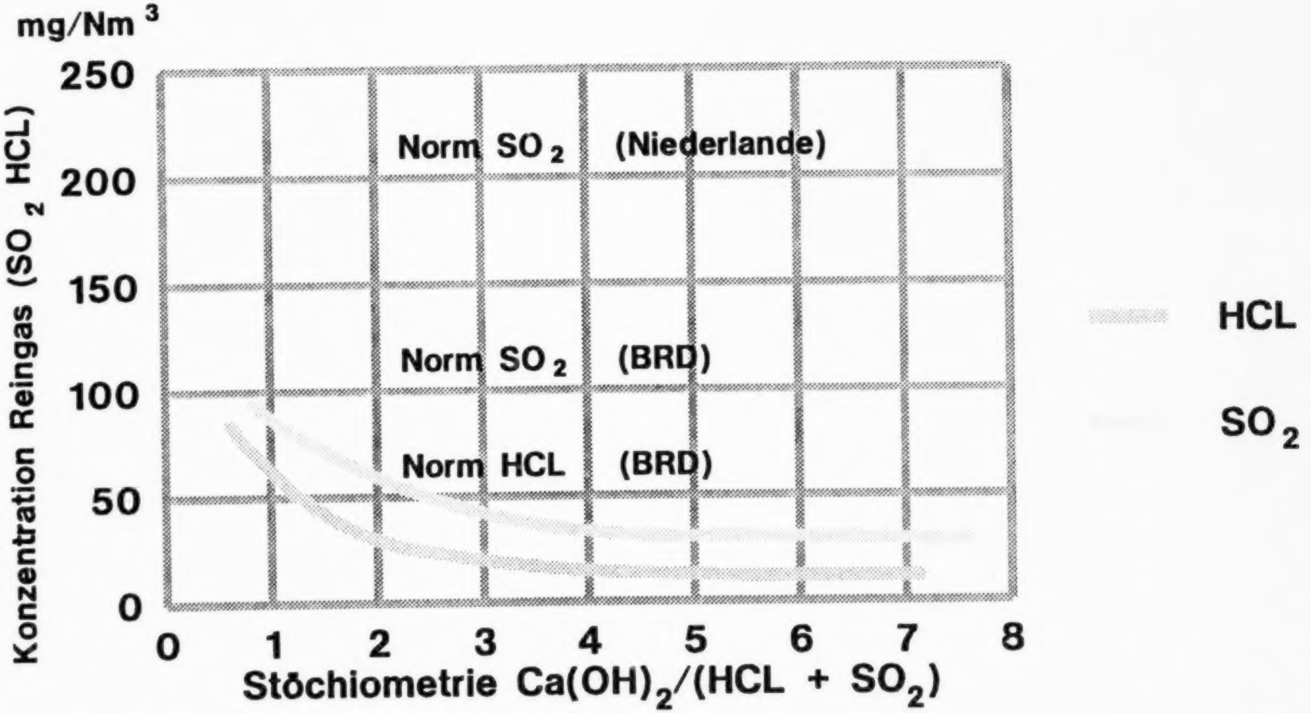


Figure 8 Absolute Konzentration von HCL und SO_2 (bezogen auf m^3 n, trocken) im gereinigten Rauchgas als Funktion des stöchiometrischen Faktors. Die Normen für HCL und SO_2 sind ebenfalls (als horizontale Linien) angegeben.

Wassergehalt im Rauchgas	ca. 14%
Reaktionstemperatur:	ca. 130 °C
Bezugssauerstoffgehalt	11%
Sauerstoffgehalt, gemessen	12 - 15%

TABLE I. VARIETY OF BOILER/FEEDER TYPES

Boiler/feeder	: chaingrate, screwfeeder, stepped grate, vertical spreader stoker, pulverized fuel, fluid bed.
Boiler types	: hot water, steam - 13 bar. (390 PSIG)
Capacities	: 0.5 - 15 MW.
Flue gas temperatures	: 90 - 260 deg. C. (194 deg. F. - 500 deg. F.)
Fuels	: coal, lignite, wood, RDF, animal fat.
Filter sizes	: 30 - 800 m2. 300 - 8000 sqft filter area.

TABLE II. FILTER FABRICS

Media	Max. operating temperature
Dralon T (PAC)	: 130 deg. C. - 270 deg. F.
Glass felt	: 260 deg. C. - 500 deg. F.
Nomex acid resist treated	: 200 deg. C. - 400 deg. F.
Ryton/Ryton, Ryton/Rastex	: 180 deg. C. - 360 deg. F.
Teflon/Teflon scrim	: 260 deg. C. - 500 deg. F.
PTFE/Rastex scrim	: 260 deg. C. - 500 deg. F.
Tefaire	: 260 deg. C. - 500 deg. F.

TABLE III. FABRIC OPERATING EXPERIENCE

Media	Operating temp. range	Life	Remarks
Poly Acrylnitrile (Dralon T)	max. 130 deg.C. 270 deg.F.	reasonable to good	some pressure drop problems premature attack. low cost, 1-3 years life.
Fibre glass felt	max. 260 deg.C. 500 deg.F.	poor	shortlife, caused by acid condensation
Aromatic Polyamide (Nomex)	max. 200 deg.C. 392 deg.F.	disappointing in some places	good at const. operating temperature above acid dew point and/or with lime injection. Good with lignite.
Ry/Rastex	max. 180 deg.C. 360 deg.F.	good	relative high cost.
Ry/Ry	max. 180 deg.C. 360 deg.F.	good critical with bromine	acceptable cost.
Teflon/Teflon	max. 260 deg.C. 500 deg.F.	good	high cost, low filtration velocity.
PTFE/PTFE	max. 260 deg.C. 500 deg.F.	good	same
Tefaire	max. 260 deg.C. 500 deg.F.	good	very good potential as a result of low Δp at high air to cloth ratio's.

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